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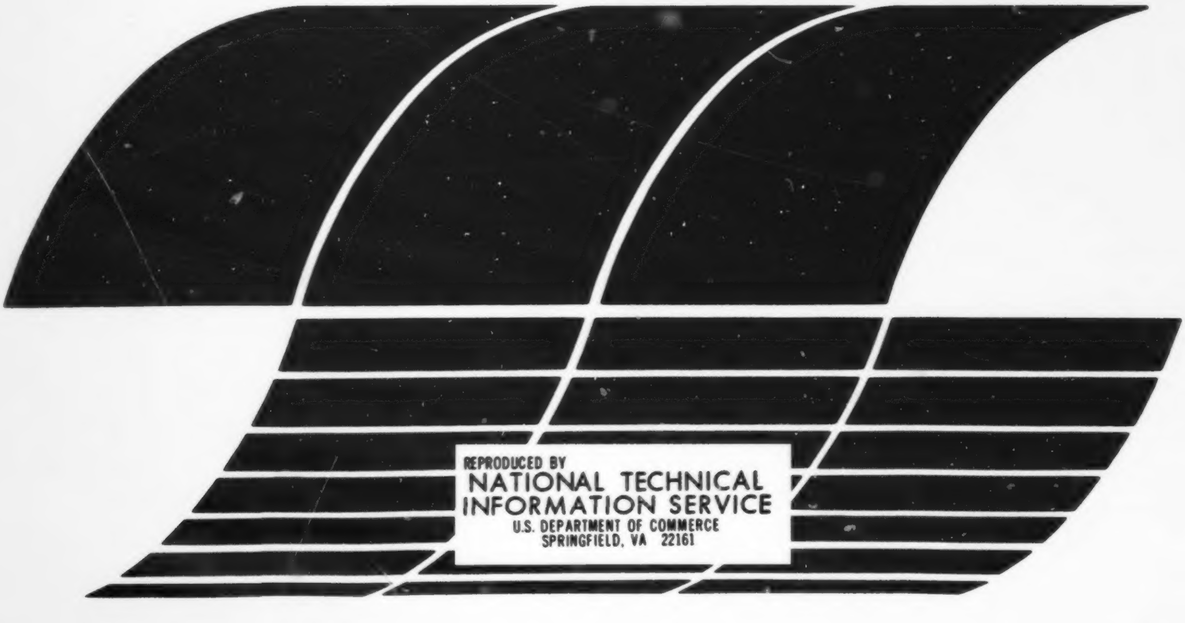

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16. ABSTRACT The report gives results of a preliminary economic evaluation of seven flue gas treatment (FGT) processes for the removal of NOx from power plant flue gas. The base case was a new, 500-MW power plant burning 3.5% sulfur coal and emitting 600 ppm NOx in the flue gas. Total capital investments and annual revenue requirements for three dry NOx-only removal processes were $38-48/kW and 2.1-3.6 mills/kWh, respectively. A dry process to simultaneously remove both SOx and NOx had a capital cost of $139/kW and an annual revenue requirement of 6.6 mills/kWh. Wet SOx/NOx systems, which also remove particulates, had capital costs and annual revenue requirements of $205-482/kW and 12.2-19.8 mills/kWh, respectively. To permit comparisons between the alternative NOx control systems, the costs of both a limestone slurry flue gas desulfurization unit and an ESP were added to the cost of the dry NOx-only processes, and the cost of an ESP was added to the cost of the dry SOx/NOx processes. Total capital investments for these combined systems based on dry NOx removal (both NOx-only and SOx/NOx removal) were about $165-175/kW. Unit revenue requirements for the combined systems were 7.1-8.6 mills/kWh for the dry processes. For the base case, wet FGT processes do not appear to be economically competitive with either the dry NOx-only or SOx/NOx systems.

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ABSTRACT

A preliminary economic evaluation was made to compare seven flue gas treatment (FGT) processes for the removal of nitrogen oxides (NO_x) from power plant flue gas. The base case power plant was a new, 500-MW unit burning 3.5% sulfur coal and emitting 600 ppm NO_x in the flue gas. The capital investments and annual revenue requirements for three dry NO_x-only removal processes ranged from $38/kW to $48/kW, and 2.1 mills/kWh to 3.6 mills/kWh respectively. A dry process for simultaneous removal of both sulfur oxides (SO_x) and NO_x had a capital cost of $139/kW and an annual revenue requirement of 6.6 mills/kWh. Wet SO_x-NO_x systems, which also removed particulates, had capital costs and annual revenue requirements ranging from $205/kW to $482/kW and 12.2 mills/kWh to 19.8 mills/kWh respectively. To permit comparisons between the NO_x control systems, the costs of both a limestone slurry flue gas desulfurization (FGD) unit and an electrostatic precipitator (ESP) were added to the cost of the dry NO_x-only processes and the cost of an ESP was added to the cost of the dry SO_x-NO_x process. The capital investments for these combined systems based on dry NO_x removal (both NO_x-only and SO_x-NO_x removal) were about $165-$175/kW. The unit revenue requirements for the combined systems ranged from 7.1 to 8.6 mills/kWh for the dry processes. Thus it would appear that for the base case power unit, the wet FGT processes are not economically competitive with either the dry NO_x-only or the dry SO_x-NO_x systems.

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ABBREVIATIONS AND CONVERSION FACTORS

ABBREVIATIONS

abs	absolute	kW	kilowatt
aft^3	actual cubic feet	kWh	kilowatthour
bbl	barrel	l	liter
Btu	British thermal unit	lb	pound
$^{\circ}$C	degrees Celsius	m	meter
cm	centimeter	M	million (mega, = 10^6)
dia	diameter	mg	milligram
ESP	electrostatic precipitator	min	minute
$^{\circ}$F	degrees Fahrenheit	mm	millimeter
FGD	flue gas desulfurization	mol	mole
FGT	flue gas treatment	MW	megawatt
ft	feet	Nm^3	normal cubic meter
ft/sec	feet per second	ppm	parts per million
g	gram	psi	pounds per square inch
G	billion (giga, = 10^9)	SCR	selective catalytic reduction
gal	gallon	sec	second
gpm	gallons per minute	SFGT-N	Shell Flue Gas Treatment (NO_x-only)
gr	grain	SFGT-SN	Shell Flue Gas Treatment (SO_x-NO_X)
hp	horsepower	sft^3/hr	standard cubic feet per hour
hr	hour	sft^3/min	standard cubic feet per minute
in.	inch	vol	volume
k	thousand	wt	weight
kg	kilogram	yr	year
kl	kiloliter		

CONVERSION FACTORS

EPA policy is to express all measurements in agency documents in metric units. Most values in this report are given in English units for the convenience of engineers and other scientists accustomed to using the English systems. The following conversion factors may be used to provide metric equivalents.

To convert from English units	To metric units	Multiply by
acre	hectare	0.405
barrel[a]	liters	158.97
British thermal unit	gram-calories	252
degrees Fahrenheit-32	degrees Celsius	0.5555
feet	centimeters	30.48
square feet	square meters	0.0929
cubic feet	cubic meters	0.02832
feet per minute	centimeters per second	0.508
cubic feet per minute	cubic meters per second	0.000472
gallons	liters	3.785
gallons per minute	liters per second	0.06308
gallons per thousand actual cubic feet (50°C)	liters per normal cubic meter	0.1608
grains	grams	0.0648
grains per cubic foot	grams per cubic meters	2.288
horsepower	kilowatts	0.746
inches	centimeters	2.54
inches of water	Pascals (Newton per square meter)	249.09
pounds	kilograms	0.4536
pounds per square inch	Pascals	6895
standard cubic feet per minute (60°F)	normal cubic meters per hour (0°C)	1.6077
tons per hour	kilograms per second	0.252

a. Forty-two U.S. gallons per barrel.

Other conversion factors to convert from	To	Multiply by
dollars	yen (Spring 1978)	220
megawatt (electric)	normal cubic meters per hour	3000
megawatt (electric)	standard cubic feet per minute	1870

PRELIMINARY ECONOMIC ANALYSIS OF NO_x FLUE GAS TREATMENT PROCESSES

EXECUTIVE SUMMARY

The existing primary and secondary annual ambient air-quality standard for nitrogen dioxide (NO_2) of 100 $\mu g/m^3$ is being met in the United States except for a few locations. However, the 1977 Clean Air Act Amendments require the promulgation of a short-term NO_2 ambient air-quality standard (3 hours or less averaging time) unless it can be demonstrated that the standard is unnecessary for public health protection. This short-term ambient standard may require additional control of nitrogen oxides (NO_x) from stationary sources, including electric power generating facilities. The 1977 Clean Air Act Amendments also require the installation of the Best Available Control Technology (BACT) for NO_x control for new or modified facilities, including electric generating facilities. Finally, Prevention of Significant Deterioration (PSD) regulations for NO_2 are also required by the 1977 Clean Air Act Amendments. These requirements of the 1977 Clean Air Act Amendments may lead to more need for NO_x emission control for stationary sources, including electric power generating facilities in the future.

For the present time, combustion modification techniques under investigation by the Electric Power Research Institute (EPRI) and the U.S. Environmental Protection Agency (EPA) may provide sufficient control of power plant NO_x emissions. However, the NO_x control capabilities of these techniques may not be adequate to meet the requirements of future regulations and alternative approaches, though potentially less cost-effective, cannot be ruled out. Therefore, NO_x flue gas treatment (FGT) processes warrant further investigation.

In a previous report covering the first phase of a multiphase NO_x removal FGT process study, a state-of-the-art survey was made of all known NO_x FGT processes undergoing development in the United States and Japan (Technical Assessment of NO_x Removal Processes for Utility Application, EPA-600/7-77-127, EPRI AF-568, TVA Y-120). Because low NO_x emission limits have been emphasized in Japan in recent years, NO_x FGT technology is more advanced in that country and most of the processes included in the first phase of the NO_x process study were developed by Japanese companies. During the preparation of that report, various Japanese and American companies were contacted for pertinent technical and economic information outlining their NO_x removal processes. This information and data obtained from other sources were combined to describe the major technical and economic aspects of each process. The description for each process consisted of a process description, status of development, reported economics, utility and raw material requirements, technical and environmental considerations, and the relative advantages and disadvantages. A major objective of the first phase was to determine candidates for preliminary economic analysis as a second-phase study.

In this report covering the second phase of the NO_x process study, the preliminary economics, comprising total capital investment and annual revenue requirement estimates, were evaluated for the selected NO_x FGT processes. The economics were calculated based on a consistent set of design and economic premises that have formed the basis for many previous flue gas desulfurization (FGD) studies done by the Tennessee Valley Authority (TVA). Although various conclusions have been advanced concerning NO_x FGT processes, no rigorous economic comparisons have been made. The primary purpose of this study is to compare the process economics of the different types of NO_x control technology on a consistent basis.

Current NO_x FGT technology involves many types of processes in different, generally early, stages of development. Selection of processes for this evaluation was therefore based both on technical and economic assessment and on the need to evaluate different types of processes. All the processes evaluated in this study are designed for 90% NO_x removal but most are in an early stage of development and none have been extensively demonstrated with coal-fired flue gas.

After the publication of the first-phase report, work was terminated by the developer of one of the eight processes chosen to undergo preliminary economic analysis and thus the process was omitted from this report (MON Alkali process). The remaining seven NO_x FGT processes are listed in Table S-1 according to the specific type of FGT process they represent--wet sulfur oxides (SO_x)-NO_x, dry SO_x-NO_x, or dry NO_x-only.

The Asahi process uses a sodium sulfite (Na_2SO_3) solution to absorb both SO_x and NO_x. The NO_x is directly absorbed with the aid of a ferrous-chelating compound. After absorption, the NO_x is eventually reduced to molecular nitrogen and vented to the atmosphere. The sulfur dioxide (SO_2) is eventually removed as byproduct gypsum ($CaSO_4 \cdot 2H_2O$) and sodium sulfate (Na_2SO_4).

The process developed by Ishikawajima-Harima Heavy Industries (IHI) uses ozone injected into the flue gas stream to oxidize nitric oxide (NO) to NO_2. The NO_2 and SO_x are absorbed from the flue gas into a calcium-based slurry. The NO_x and SO_x are ultimately removed as molecular nitrogen and byproduct gypsum, respectively.

TABLE S-1. NO_x FGT PROCESSES SELECTED FOR EVALUATION

Process	Type
Wet SO_x-NO_x	
Asahi	Absorption-reduction
Ishikawajima-Harima Heavy Industries (IHI)	Oxidation-absorption-reduction
Moretana Calcium	Oxidation-absorption-reduction
Dry SO_x-NO_x	
UOP Shell Flue Gas Treatment (SFGT-SN)	Sorption of SO_x and selective catalytic reduction of NO_x
Dry NO_x-only	
Hitachi Zosen	Selective catalytic reduction
Kurabo Knorca	Selective catalytic reduction
UOP Shell Flue Gas Treatment (SFGT-N)	Selective catalytic reduction

In the Moretana Calcium process, NO is oxidized to NO_2 by injecting chlorine dioxide (ClO_2) into the flue gas stream. The NO_2 and SO_x are absorbed into a limestone slurry in a Moretana plate tower absorber. The absorbed NO_2 and SO_x form various calcium salts. The SO_x is ultimately discarded as a mixed sulfite-sulfate sludge. NO_x is removed as ammonium nitrate (NH_4NO_3) after reaction of the calcium nitrates [$Ca(NO_3)_2$] with ammonium sulfate [$(NH_4)_2SO_4$].

The Universal Oil Products, Inc., (UOP) Shell Flue Gas Treatment (SFGT) process for simultaneous SO_x-NO_x removal (SFGT-SN) uses a copper oxide (CuO) acceptor for SO_x removal. As the SO_x is absorbed, the CuO is converted to copper sulfate ($CuSO_4$). The sulfate is the catalyst for the selective catalytic reduction (SCR) of NO_x to molecular nitrogen and water by reaction with ammonia (NH_3). Subsequent regeneration of the multiple catalyst beds produces a SO_2-rich stream which can be used to produce various byproducts, e.g., sulfur or sulfuric acid (H_2SO_4). This process uses a parallel-flow, fixed-bed reactor located between the boiler economizer and the boiler air heater.

The three remaining processes control only NO_x. Each process is based on SCR, i.e., NH_3 selectively reacts with NO_x in the presence of a catalyst to form molecular nitrogen and water. The reactors for each of these processes are located between the boiler economizer and the boiler air heater. These processes are identified as follows:

1. Hitachi Zosen process: A fixed-bed reactor is used in which the flue gas passes in parallel flow along the surface of the corrugated-shaped catalyst cells.

2. Kurabo Knorca process: This process uses a moving-bed reactor. The flue gas stream is passed in crossflow through the bed of spherically shaped and slowly moving catalyst. This reactor must be located downstream from a hot electrostatic precipitator (ESP).

3. UOP SFGT-N process: A parallel passage, fixed-bed reactor is used in which the catalyst is contained in unit cells and the flue gas is forced across the face of the catalyst layer.

Since some of these FGT processes, as offered by the various vendors, remove other potential air pollutants in addition to NO_x, the following terminology and equipment additions have been assumed to permit the comparison of NO_x FGT processes on a consistent basis.

1. Wet SO_x-NO_x: These processes simultaneously remove both SO_x and NO_x by absorption into an aqueous solution. Particulate collection is an inherent part of these processes and is achieved by either a prescrubber or an ESP and a prescrubber in series. These processes included all those listed as wet SO_x-NO_x in Table S-1.

2. Dry SO_x-NO_x: This process simultaneously removes both SO_x and NO_x. SO_x removal is achieved by using a dry sorbent while NO_x removal is by SCR. In all economic comparisons an ESP has been included for particulate matter (PM) control.

3. SCR-FGD: These are the dry, SCR-based, NO_x-only processes previously identified in Table S-1. For all economic comparisons an ESP and a limestone slurry process have been included for PM and SO_2 control.

Thus the following comparisons can be made on a consistent set of design and economic premises.

1. Wet, oxidation-absorption-reduction versus wet, absorption-reduction.

2. Wet SO_x-NO_x versus dry SO_x-NO_x.

3. Wet SO_x-NO_x versus SCR-FGD.

4. Dry SO_x-NO_x versus SCR-FGD.

5. Moving-bed reactor versus parallel flow, fixed-bed reactor for SCR.

DESIGN AND ECONOMIC PREMISES

A specific set of design and economic premises was established for the comparative calculations. The basic design and economic premises used in this report were established by TVA and EPA for power plant stack gas emission control studies. Although these premises were originally designed for FGD studies, appropriate additions and minor adjustments (primarily dealing with NO_x control), after joint consultation between TVA, EPA, and EPRI, were made and the resulting premises were used for this study.

Design Premises

The power plant assumed as a basis for this study was a new, 500-MW coal-fired boiler at a midwestern location. The coal had a heating value of 10,500 Btu/lb and contained 3.5% sulfur and 16% ash. The coal feed rate of 428,600 lb/hr was based on a heat rate of 9,000 Btu/kWh. On-stream time for the boiler was 7,000 hr/yr.

The FGT system included all equipment (based on current technology and information) necessary to achieve 90% NO_x removal from flue gas containing 600 ppm NO_x and to convert this NO_x to environmentally inert substances. Other design conditions included a minimum stack gas temperature of 175°F, assumed proven technology (only pumps are spared), and solids disposal ponds or landfills located one mile from the FGT facility. When reheat was required an indirect steam reheat system was included.

The wet FGT processes were equipped with a common inlet plenum. This plenum equally distributed the flue gas from the boiler air heaters to each of the processing trains. However, the dry FGT processes operate at higher temperatures (572°F-752°F) and for new boilers, as used in this study, the FGT system was installed between the economizer and the boiler air heater. (For retrofit applications the flue gas may have to be taken from the air heater and reheated before entering the FGT system.)

In addition to preparing the design premises for the NO_x FGT process itself, the design premises for an overall FGT system, including PM removal and FGD, were included to allow for comparisons of the various dry and wet processes. Because the wet SO_x-NO_x processes typically use a prescrubber for PM removal and achieve greater than 90% SO_x removal, the design premises for the FGT system included the following removal efficiencies.

1. Particulate: 99.5%
2. SO_x: 90%
3. NO_x: 90%

For FGT system comparisons, a high-efficiency ESP and a 90% efficient limestone "scrubbing" FGD system were included with the NO_x-only FGT processes and a high-efficiency ESP was included with the dry SO_x-NO_x

FGT process. The economics of the limestone FGD system have been previously calculated based on identical TVA premises. Thus the resulting capital investments and annual revenue requirements for FGD and ESP were added to the NO_x-only FGT processes in calculations of the economics of the total package system.

The costs associated with PM disposal are not included, but would be essentially the same for each process. However, the costs associated with fly ash handling are included. Also for the dry processes, the costs which may be required for additional soot blowing or washing of the air heater are excluded in this estimate. Included in the estimate of the dry SO_x-NO_x process are the incremental cost differences which are required for the air heater to recover vendor-claimed heat credits.

Economic Premises

Capital investment estimates were based on mid-1979 construction costs. Current equipment costs were obtained from in-house information or from equipment vendors. These costs were converted to 1979 dollars by assuming an escalation rate of 8% annually. The equipment installation charges were calculated as percentages of the equipment cost determined by previous TVA experience. Indirect investments such as engineering design and supervision and architect and engineering expense were calculated as percentages of the equipment cost. Indirect investments such as construction expense and contractor fees were calculated as a percentage of the total direct investment. Royalty fees were included.

Annual revenue requirements were based on mid-1980 operating costs using average capital charges with a 7,000-hr/yr on-stream time. The current delivered costs of raw materials (obtained from suppliers in the Chicago area) were escalated by 5%-10% per year, depending on the expectations of these suppliers. The other annual operating charges were estimated based on ranges obtained from standard chemical engineering sources and on previous TVA experience. Byproduct sales credits were given only for 98% H_2SO_4, based on the projected 1980 sales price of this byproduct material. Spent catalyst disposal costs or credits were applied based on disposal cost or scrap metal value of the base catalyst support material.

PROCESS DESCRIPTIONS AND SYSTEMS ESTIMATED

As expected, both wet and dry SO_x-NO_x processes were more complicated than the NO_x-only processes. Because the NO_x-only processes convert the NO_x to inert molecular nitrogen by contacting NO_x with NH_3 in the presence of a catalyst, these processes have only two, relatively small, processing sections: an NH_3 storage and injection system and a catalytic reactor. The dry NO_x-only processes in this study are the UOP SFGT-N, Kurabo, and Hitachi Zosen processes.

The UOP SFGT-SN process provides simultaneous SO_x-NO_x control. This UOP process is unique in that the sorbent for SO_2 forms a base-metal sulfate which in combination with NH_3 reduces NO_x to molecular nitrogen. Subsequent regeneration of the multiple catalyst beds using hydrogen gas from a steam-naphtha reformer yields a concentrated SO_2 stream which can be converted to H_2SO_4 in an onsite plant. However, there are alternatives for hydrogen production, such as coal gasification, and for SO_2 processing, such as elemental sulfur production with a Claus unit.

The wet SO_x-NO_x processes absorb both SO_x and NO_x in an aqueous solution to form sulfite-sulfate salts and differing amounts of both nitrate salts and complex nitrogen-sulfur compounds. The sulfite-sulfate salts are typically converted into byproduct gypsum and the complex nitrogen-sulfur compounds are thermally decomposed into molecular nitrogen and sulfate salts. The nitrate salts are either decomposed with the complex nitrogen-sulfur compounds or they are concentrated in a byproduct stream to form a fertilizer material. The wet SO_x-NO_x processes included in this study are the Asahi, IHI, and Moretana Calcium processes.

Process descriptions, detailed flowsheets, material balances, and a current status of development summary were prepared for each process based on data supplied by the process vendors. Equipment descriptions and costs were prepared from the available data for each process.

ECONOMIC EVALUATION AND COMPARISON

The process economics consist of capital investment and annual revenue requirements for each system. The economic estimates were based on the material balances and equipment lists, and followed standard engineering practices. Because of the eclectic nature of the source data, the simplifying assumptions made, and the necessity of projecting cost into the future, these preliminary estimates are considered to be accurate to an overall variation of -20% to +40%.

The coal-fired unit has an output of 500 MW including all system energy requirements up to and including the ESP and the induced-draft fans. Thus the 500-MW rating does not include the energy usage of the FGT or FGD processes. However, throughout this report, for the SO_x-NO_x-PM control systems, the capital investment and annual revenue requirement are given in dollars per kW and mills per kWh based on a net MW rating. This net MW rating is calculated by derating the 500-MW output for the energy usage of the FGT and FGD processes.

Comparison of capital estimates of the FGT processes between any two reports is very difficult on a dollar per kW or absolute dollar basis. The FGT processes are primarily sized on flue gas flow rate and this flow rate may vary significantly even between studies for plants of similar MW rating. This may be due to several reasons such as differences in coal composition, coal heating value, plant heat rate, usage of gross versus net MW rating, etc. To reduce this difficulty of comparison between reports, the capital investments of the SO_x-NO_x-PM control systems are also shown on a $\$/aft^3/min$ basis. The flue gas flow rate is 2.146 $Maft^3/min$ at the economizer outlet at 705°F.

Capital Investment Estimates

Capital investment is the sum of direct investments, indirect investments, contingencies, land costs, working capital, and royalty fees. The total capital investment estimates for the seven NO_x FGT processes are shown in Table S-2. Where applicable, the capital investments for the limestone FGD system and ESP are shown as well as the resulting capital investment for the combined NO_x, SO_x, and PM removal systems.

TABLE S-2. CAPITAL INVESTMENT FOR

THE BASE CASE SO_x-NO_x-PM SYSTEMS

	Total capital investment					
	M$				$/kW	$/aft^3/min
Process	FGT	FGD	ESP	Total	Total	Total
UOP SFGT-N	18.4	50.4	10.8	79.6	165	37.1
UOP SFGT-SN[a]	67.2	-	14.6	81.8	169	38.1
Kurabo Knorca	21.2	50.4	12.1	83.7	174	39.0
Hitachi Zosen	23.3	50.4	10.8	84.5	175	39.4
Moretana Calcium	88.0	-	7.2	95.2	205	44.4
Asahi	104.9	-	-	104.9	233	48.9
IHI	203.6	-	-	203.6	482	94.9

a. Based on hydrogen production from naphtha and H_2SO_4 byproduct from SO_2.

The capital investments for the various NO_x-only FGT processes range from \$18M (\$38/kW) to \$23M (\$48/kW). For the combined SCR-FGD systems, the capital investments range from \$80M (\$165/kW) to \$85M (\$175/kW). For the SO_x-NO_x processes the capital investments range from \$67M (\$139/kW) to \$204M (\$482/kW).

The capital investment for the SCR-FGD systems range from \$165/kW for the UOP SFGT-N - limestone FGD system to \$175/kW for the Hitachi Zosen - limestone FGD system. Given the uncertainties involved in this estimate it is impossible to select the best combination based on the capital investment. The capital investment for the UOP SFGT-SN system is \$169/kW, which is between the \$165/kW to \$175/kW range for the SCR-FGD systems. Based on this data, it is impossible to select between the SCR-FGD and UOP SFGT-SN systems.

Among the wet processes, the IHI process, with a capital investment of \$482/kW, is not economically attractive under the conditions studied. The Moretana Calcium process, with a capital investment of \$205/kW, is about 24% more costly than the least expensive SCR-FGD system and 21% more costly than the dry SO_x-NO_x system. The capital investment for the Asahi process is \$233/kW, which is about 40% higher than the least expensive SCR-FGD system and 38% higher than the dry SO_x-NO_x system. Though the capital investments are higher for the Asahi and Moretana Calcium processes, neither can be eliminated on the basis of capital investment alone.

Annual Revenue Requirement Estimates

The annual revenue requirements consist of direct costs for raw materials, conversion and services, and indirect costs for capital charges and overheads. The annual revenue requirements for the seven NO_x FGT processes are shown in Table S-3. The total annual revenue requirements for the limestone FGD process and the ESP as well as the resulting annual revenue requirements for the combined system are also shown for those NO_x FGT processes where an additional FGD process and/or an ESP is needed to allow a direct comparison. In addition, the unit revenue requirements in mills/kWh for each system are shown.

TABLE S-3. ANNUAL REVENUE REQUIREMENTS FOR

THE BASE CASE SO_x-NO_x-PM SYSTEMS

Process	Annual revenue requirements, M\$ FGT	FGD	ESP	Total	Equivalent unit revenue requirements, mills/kWh Total
UOP SFGT-N	7.2	14.7	2.2	24.1	7.13
UOP SFGT-SN[a]	22.5	-	3.0	25.5	7.53
Kurabo Knorca	9.3	14.7	2.7	26.7	7.91
Hitachi Zosen	12.2	14.7	2.2	29.1	8.60
Moretana Calcium	38.1	-	1.5	39.6	12.20
Asahi	39.8	-	-	39.8	12.63
IHI	58.6	-	-	58.6	19.82

a. Based on hydrogen production from naphtha and H_2SO_4 byproduct from SO_2.

The annual revenue requirements show a much wider spread in values for the various processes than was apparent in the total capital investment estimates. These annual revenue requirements range from \$7.2M for the UOP SFGT-N process to \$58.6M for the IHI process. When the costs for the limestone FGD process and ESP are added in order to compare total package systems, the annual revenue requirements increase to \$24.1M for the UOP SFGT-N - limestone FGD system.

Based on the unit revenue requirements, the wet SO_x-NO_x processes do not appear as economically attractive as dry processes for the conditions studied. When compared with the least expensive SCR-FGD system, the unit revenue requirements for the wet SO_x-NO_x systems are about 71%, 77%, and 178% higher for the Moretana Calcium, Asahi, and IHI systems respectively. These higher unit revenue requirements indicate that the wet SO_x-NO_x systems are not economically competitive with the dry NO_x control systems under the conditions used in this evaluation.

The UOP SFGT-SN process, with the credit for the sale of byproduct H_2SO_4 at \$30/ton, has a unit revenue requirement of 7.53 mills/kWh. Without this credit the revenue requirements is about 14% higher.

In a comparison of the wet SO_x-NO_x processes, the Asahi process, which is based on the direct absorption of NO_x, and the Moretana Calcium process, which uses ClO_2 as a gas-phase oxidant, have similar unit revenue requirements. The unit revenue requirement is significantly higher for the IHI process, which uses ozone as the gas-phase oxidant, than for the Asahi or Moretana Calcium processes.

Based on an accuracy range of -20% to +40%, the resulting range of estimated capital investment and annual revenue requirement for the SO_x-NO_x-PM control systems are given in Table S-4. These capital investment and annual revenue requirement ranges are also displayed in Figures S-1 and S-2, respectively.

TABLE S-4. ACCURACY RANGE OF THE ESTIMATED CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENT FOR THE SO_x-NO_x-PM SYSTEMS

(Based on a -20% to +40% range)

Process	Capital investment, \$/kW			Unit revenue requirements, mills/kW		
	Low	Base	High	Low	Base	High
UOP SFGT-N	132	165	231	5.70	7.13	9.98
UOP SFGT-SN[a]	135	169	237	6.02	7.53	10.54
Kurabo Knorca	139	174	244	6.33	7.91	11.07
Hitachi Zosen	140	175	245	6.88	8.60	12.04
Moretana Calcium	164	205	287	9.76	12.20	17.08
Asahi	186	233	326	10.10	12.63	17.68
IHI	386	482	675	15.86	19.82	27.75

a. Based on hydrogen production from naphtha and H_2SO_4 byproduct from SO_2.

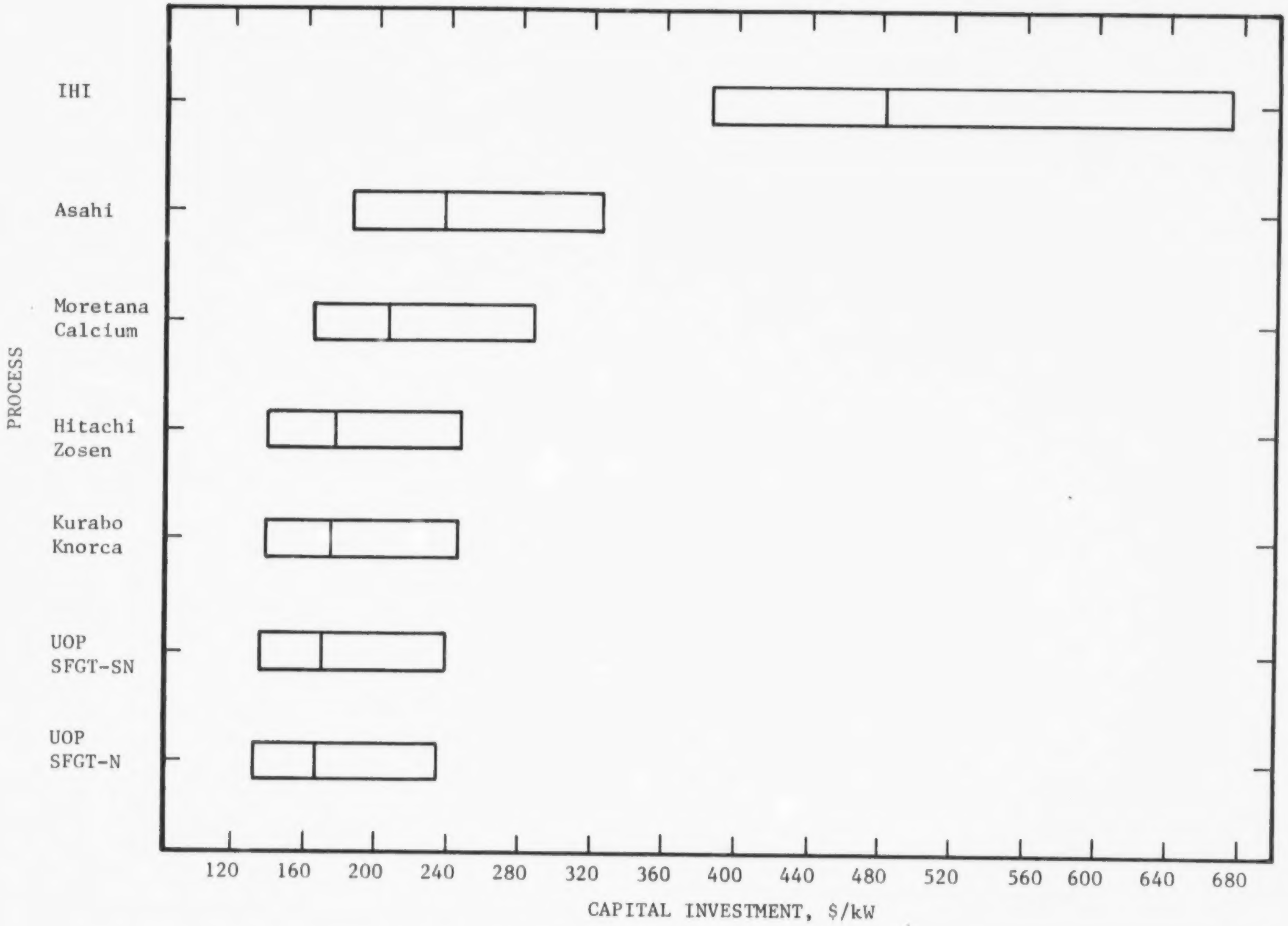

Figure S-1. Accuracy range of the estimated capital investment for the alternative SO_x-NO_x-PM systems (based on a -20% to +40% range).

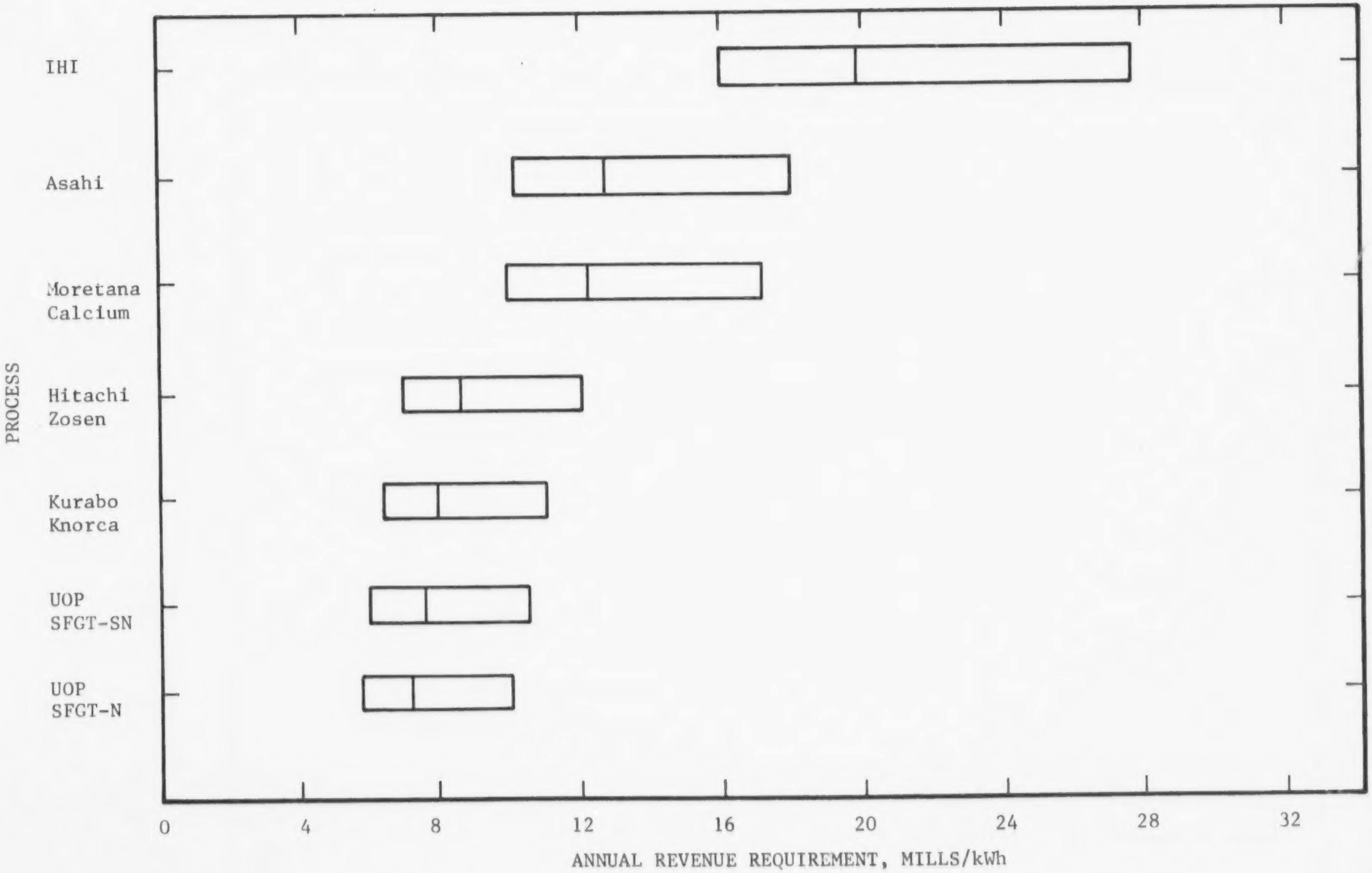

Figure S-2. Accuracy range of the estimated annual revenue requirement for the alternative SO_x-NO_x-PM systems (based on -20% to +40% range).

The results of these comparisons are summarized in Table S-5. The wet SO_x-NO_x processes do not appear economically attractive for new power plant applications when compared with either the dry SO_x-NO_x process or the SCR-FGD systems. Comparisons between the dry UOP SFGT-SN process and SCR-FGD systems are inconclusive. Comparisons within the SCR-FGD systems, i.e., moving versus parallel flow, fixed-bed reactors are also inconclusive. However, recent trends in Japan indicate the fixed-bed reactor systems are preferred.

Energy Consumption

Because of the likelihood of future increases in the costs of energy, the energy consumption of FGT methods has become increasingly important. The energy requirements for each of the seven NO_x FGT systems, the ESP's, and the limestone FGD unit were estimated and are shown in Table S-6. The fuel, steam, electricity, and heat credits are shown as equivalent MBtu/hr. (The electricity is also shown as equivalent megawatts.) The total energy consumption for each system is shown as a percent of the boiler capacity.

The dry processes consume the least net energy, 4% of the boiler capacity. As a result of large heat credits claimed by UOP, the UOP SFGT-SN process has a slightly lower net energy use, 3.6% of boiler capacity, than the SCR-FGD systems, where energy use ranges from 3.7% to 3.9% of boiler capacity. These heat credits claimed by UOP consist of the following: (1) heat of compression from work done on the flue gas by the flue gas blower, (2) heat of reaction released during acceptance and regeneration, (3) heat of combustion released to flue gas from purged regeneration gas remaining in the reactor at the end of regeneration, (4) heat of combustion from burning of excess reducing gas, and (5) extra heat available from the flue gas resulting from removal of sulfur trioxide (SO_3) and subsequent lowering of the dewpoint temperature. The heat required from the flue gas and regeneration gas for heating steam and the NH_3 air mixture was debited to the UOP SFGT-SN process. The actual recovery of portions of the heat credits claimed by UOP to be recovered at the air heater may require additional capital expenditures. The incremental capital investment for a larger air heater to recover these heat credits has been included in this report. The IHI process consumes the most energy (about 19%). The order by process type in increasing energy costs is dry SO_x-NO_x, SCR-FGD, and wet SO_x-NO_x. The major energy consumption is electricity for all of the SO_x-NO_x processes except the UOP SFGT-SN process, in which fuel is the largest energy consumption. In most of the processes, equipment drives, particularly for the flue gas blowers, consume most of the electricity. However, for the IHI process, most of the electricity is consumed in generating ozone. Heat credits from the exothermic NO_x reduction reactions recoverable at the air heater have been given to the dry processes and hence subtracting these from the fuel, steam, and electrical requirements give the total energy consumption shown in the table. Although the UOP SFGT-SN process required incremental air heater costs to recover

TABLE S-5. RESULTS OF ALTERNATIVE SO_x-NO_x COMPARISONS

Alternatives	The more attractive alternative	Magnitude of the difference
Wet absorption-reduction versus wet oxidation-absorption-reduction (ozone)	Wet absorption-reduction	Significant
Wet absorption-reduction versus wet oxidation-absorption-reduction (ClO_2)	Inconclusive	Insignificant
Wet SO_x-NO_x versus dry SO_x-NO_x	Dry SO_x-NO_x	Significant
Wet SO_x-NO_x versus SCR-FGD	SCR-FGD	Significant
Dry SO_x-NO_x versus SCR-FGD	Inconclusive	Insignificant
Moving-bed reactor versus parallel flow, fixed bed reactor with SCR-FGD	Inconclusive	Insignificant

TABLE S-6. COMPARISON OF ENERGY REQUIREMENTS FOR THE VARIOUS NO_x FGT PROCESSES[a]

Process	Fuel, MBtu/hr	Steam, MBtu/hr	Electricity, MBtu/hr	Heat credit, MBtu/hr	Total equivalent energy consumption,[b] % of boiler capacity
Limestone FGD	-	78	74	-	3
Cold ESP (99.5% efficient)	-	-	8	-	0.2
Cold ESP (99.5% efficient: UOP SFGT-SN	-	-	13	-	0.3
Cold ESP (94% efficient)	-	-	4	-	0.1
Hot ESP (98% efficient)	-	-	19	-	0.4
Asahi	216	105	189	-	11
IHI	23	232	581	-	19
Moretana Calcium	-	117	244	-	8
Moretana Calcium and ESP (94%)	-	117	248	-	8
Hitachi Zosen	-	1	20	14	0.2
Hitachi Zosen and FGD and ESP (99.5%)	-	79	102	14	4
Kurabo Knorca	-	1	20	16	0.1
Kurabo Knorca and FGD and ESP (98%)	-	79	113	16	4
UOP SFGT-N	-	1	20	15	0.2
UOP SFGT-N and FGD and ESP (99.5%)	-	79	102	15	4
UOP SFGT-SN	200	18	91	160	3
UOP SFGT-SN and ESP (99.5%)	200	18	104	160	4

a. Does not include energy requirement represented by raw materials.

b. Based on a 500-MW boiler, a gross heat rate of 9,000 Btu/kWh for generation of electricity, and a boiler efficiency of 90% for generation of steam.

heat credits, the SCR-FGD systems do not have the incremental costs since the heat credit from just the exothermic NO_x reduction reaction is very small. Energy requirements for the purchased raw material and PM disposal are not included.

CONCLUSIONS

The following conclusions concerning current SO_x-NO_x technology are subject to numerous important constraints which must be taken into consideration in the use of these economic data. These conclusions are strictly applicable only to the base case, new, 500-MW coal-fired boilers. Under other power plant applications, particularly retrofit operations, these conclusions may not be valid. Because of the early status of development of some of these processes (particularly the wet SO_x-NO_x processes) and the lack of any long-term continuous operational testing on coal-fired boiler flue gas, the installation and operating costs could change significantly.

The first and most obvious conclusion is that NO_x FGT processes using ozone as a gas-phase oxidant are prohibitively expensive for coal-fired power units. The total capital investment and the unit revenue requirements for the IHI process using ozone are more than twice that of a comparable dry, simultaneous SO_x-NO_x system or SCR-FGD system. Ozone generation accounts for a significant portion of these higher costs (38% of the capital investment and 10% of the energy consumption is for ozone generation). Barring some unforeseen major breakthroughs in the manufacture of ozone it is highly unlikely that NO_x FGT systems based on ozone will be economically attractive in the future.

Although the other wet SO_x-NO_x processes (the Asahi process and the Moretana Calcium process) are more expensive in terms of capital investment than the dry, simultaneous SO_x-NO_x or SCR-FGD systems, the differences are about the same order of magnitude as the uncertainties surrounding the cost estimates. Thus, even though the capital investments for the Asahi process and the Moretana Calcium process are higher, neither can be eliminated on the basis of the capital investments alone. However, on the basis of unit revenue requirements, the Asahi and the Moretana Calcium processes are significantly more expensive than the least expensive dry, SCR-FGD system. Thus, these processes appear economically unattractive at this time. (It should be noted, however, than in order to achieve the minimum 90% NO_x reduction for this study, the Moretana Calcium and Asahi processes achieve 95% and 99% SO_2 removal, respectively, which is greater than the 90% SO_2 reduction achieved by the other alternatives.)

A comparison of the dry SO_x-NO_x process and the SCR-FGD systems gives inconclusive results. A more detailed comparison of the two most promising NO_x FGT technologies, dry SO_x-NO_x and dry NO_x-only FGT, based on actual operating experience is needed before the best alternative under various circumstances can be determined. To estimate the cost of the UOP SFGT-SN process, hydrogen production from naphtha and conversion of SO_2 to a H_2SO_4 byproduct are assumed. However, if in the future the supply of naphtha becomes uncertain, it may be necessary to produce hydrogen by an alternative method, for example, coal gasification. Also, if no market is available for H_2SO_4 byproduct, SO_2 processing alternatives may be used, such as elemental sulfur, or liquid SO_2 production. Inclusion of alternative processing schemes for the UOP SFGT-SN process may significantly alter the economic comparisons presented in this study.

The capital investments of the dry NO_x-only FGT processes do not differ significantly. Variations in catalyst cost account for most of the differences. The unit revenue requirements of the dry NO_x-only FGT systems vary more than the total capital investments, but have some degree of overlap within the range of uncertainty associated with preliminary economic studies. Further complicating a comparison of unit revenue requirements is the fact that the catalyst lifetime, one of the major annual costs for the dry NO_x FGT processes, is uncertain. The lifetimes estimated by most of the process vendors cannot be verified because their catalysts have not been used for NO_x removal on coal-fired boiler flue gas during a long-term continuous operation. However, with a catalyst life significantly greater than the 1 year assumed in this study (the vendors expected 2 to 3 years), the annual revenue requirement would be substantially reduced.

The capital investment and revenue requirements for the Kurabo Knorca system, with a moving-bed reactor, range between those of the UOP SFGT-N and Hitachi Zosen systems, with parallel flow, fixed-bed reactors. Therefore no significant distinction can be stated between the moving-bed and parallel flow, fixed-bed systems. However, recent trends in Japan indicate the fixed-bed systems are preferred.

The sensitivity of the dry processes to the cost of NH_3 has been the subject of some concern. In this study, the annual NH_3 cost was less than 10% of the annual revenue requirement for two out of the three dry NO_x FGT processes (Hitachi Zosen and Kurabo Knorca), and the NH_3 cost accounted for about 13% of the annual revenue requirement for the remaining dry process (UOP SFGT-N). Thus, under the premises used in this study, the annual revenue requirement may be expected to increase about 10% when the NH_3 cost is doubled.

The estimates presented in this report are based on currently available information about the various NO_x removal processes. The full impacts of the NO_x removal processes on the performance of downstream equipment (such as wet SO_2 scrubbers, air heater, ESP or baghouse, and flue gas fan) are not completely understood at this time. For example, modifications to air heater design and operation may be necessary to avoid excessive deposition of ammonium salts. Also, treatment of scrubber byproduct and blowdown streams may be necessary to prevent release of excessive quantities of ammonium salts into the environment. Cost data presented in this report may have to be revised as the information concerning these impacts becomes available from future pilot and prototype testing.

Recommendations

Additional preliminary and definitive economic analysis of NO_x control systems should be conducted. The results and conclusions of this study are based on a single base case situation and extrapolation of these results to other conditions could lead to erroneous conclusions. Also, current and future development testing will better define the FGT technology and its impact on downstream equipment, such as the air heater, FGD system, etc. For these reasons, additional economic evaluations, including both FGT and combustion modifications, should be performed. Several variations of the base case should be included to determine the sensitivity of these NO_x control technologies to design and economic variations. The case variations should include power unit capacity, sulfur level in coal, NO_x removal efficiency, raw material stoichiometry, and byproduct sales revenue.

INTRODUCTION

Manmade nitrogen oxide (NO_x) emissions are classified as stationary or mobile, depending upon their source. Each source is responsible for approximately half of the total U.S. NO_x emissions. These sources can be further divided into the various groups listed in Table 1. The estimated amounts and the percentages of the total amount of manmade NO_x released from each source for the years 1970, 1972, and 1974 are also shown. During this period the only two groups with increasing NO_x emissions were the mobile sources, particularly automobiles and trucks, and the fuel combustion group of stationary sources.

The NO_x emission limits for mobile sources were formulated in 1971 by the U.S. Environmental Protection Agency (EPA), originally for compliance by the mid-1970's. However, because of technical, economic, and other considerations, compliance with these standards for automobiles has been delayed. Thus the emphasis on decreasing manmade NO_x emissions, at least for the near future, may be shifted toward the stationary sources. Since the fuel combustion group represents 90-95% of all the emissions from stationary sources, this group appears likely to undergo more stringent regulation.

The fuel combustion group can be divided into other subgroups according to the type of fuel used and the type of operation being performed. As one would expect, the six fuel combustion sources emitting the largest quantities of NO_x are utility and industrial boilers burning coal, oil, or gas. These six sources and the estimated amount of NO_x emitted for each are listed in Table 2. The major source of NO_x in the fuel combustion group is coal-fired utility boilers, followed by oil-fired industrial boilers and oil-fired utility boilers; approximately 30% of all stationary-source NO_x (about 15% of the total manmade NO_x) is emitted by coal-fired utility boilers.

The projected increasing reliance on coal as the major fuel for the generation of electrical energy and the gradual phaseout of oil and natural gas indicate that the NO_x contribution from large stationary sources may increase in the future. In addition, the anticipated enforcement of regulations requiring the use of coal in new, large industrial boilers and the conversion of existing large boilers from oil or gas to coal will further aggravate efforts to control U.S. NO_x emissions. In fact, it has been recently estimated (12) that with a continuation of the present controls on large utility and industrial boilers, the total annual NO_x emissions will increase about 60% from an estimated 4.7 Mtons in 1975 to 7.6 Mtons in 1985. With the widespread application of combustion modifications on new stationary sources, the total amount of NO_x emitted by large boilers may increase about 40% to nearly 6.6 Mtons in 1985 (12).

TABLE 1. NO_x EMISSIONS IN THE UNITED STATES

Source	Annual NO_x emissions					
	1970		1972		1974	
	Amount, Mtons	% of total NO_x	Amount, Mtons	% of total NO_x	Amount, Mtons	% of total NO_x
Mobile						
Highway	6.9	33.8	7.9	35.6	8.1	36.0
Nonhighway	2.4	11.8	2.6	11.7	2.6	11.6
Subtotal	9.3	45.6	10.5	47.3	10.7	47.6
Stationary						
Fuel combustion	10.1	49.5	10.8	48.6	11.0	48.9
Industrial processes (noncombustion)	0.6	2.9	0.6	2.7	0.6	2.7
Solid waste	0.3	1.5	0.2	0.9	0.1	0.4
Miscellaneous	0.1	0.5	0.1	0.5	0.1	0.4
Subtotal	11.1	54.4	11.7	52.7	11.8	52.4
Total NO_x emissions	20.4	100.0	22.2	100.0	22.5	100.0

Source: (19)

TABLE 2. NO_x EMISSIONS FROM SELECTED STATIONARY SOURCES IN THE UNITED STATES IN 1972

Stationary source	NO_x emissions Annual amount, short tons	% of stationary sources	% of manmade NO_x
Utility boilers			
Coal-fired	3,853,000	30.7	15.6
Oil-fired	1,228,000	9.8	5.0
Gas-fired	920,000	7.3	3.7
Industrial boilers			
Coal-fired	810,000	6.5	3.3
Oil-fired	1,372,000	11.0	5.6
Gas-fired	541,000	4.3	2.2
Total	8,724,000	69.6	35.4

Source: (15)

Following enactment of the Clean Air Act of 1970, EPA released New Source Performance Standards (NSPS) for NO_x emissions in December 1971. The latest NSPS were released by EPA in June 1979. Both of these sets of standards for large boilers (>250 MBtu/hr, 28-MW equiv) are shown in Table 3. The most recent standards are based on nitrogen in fuels (i.e., ranging from 0.2 lb NO_x/MBtu for natural gas with no bound nitrogen to 0.6 lb NO_x/MBtu for coal which contains about 1% bound nitrogen), the size of the utility boiler, and the best available technology for NO_x control. However, the NSPS may become more stringent as NO_x control technology is further developed (see projected research objectives in Table 3). The need for additional controls may also increase as the amount of NO_x emitted to the atmosphere increases.

NO_x EMISSION CONTROL

Although NO_x is present in relatively minor amounts in the flue gas, a single, 500-MW coal-fired power plant releases approximately 10,800 tons/yr of NO_x to the atmosphere. The NO_x in the flue gas is generated during combustion from the oxidation of nitrogen-containing compounds in the fuel (fuel NO_x) and the thermal reaction of molecular nitrogen and oxygen during combustion (thermal NO_x). The portion of the total NO_x emissions from each of these sources is a function of many variables and is quite site specific.

Of the two types of NO_x control technology currently under development, combustion modifications and flue gas treatment (FGT), combustion modifications will probably receive the initial emphasis since it is considered the most cost-effective method of controlling NO_x emissions (12).

TABLE 3. NO_x EMISSION STANDARDS AND PROJECTED RESEARCH OBJECTIVES FOR LARGE FOSSIL FUEL-FIRED BOILERS[a]

	1971 EPA standard (18)		June 1979 NSPS (20)	Projected research objectives (15)
	Lb NO_x/MBtu input to boiler[b]	NO_x ppm[c]	Lb NO_x/MBtu input to boiler	1985 NO_x ppm[b]
Gaseous fuel	0.2	150	0.2	50
Liquid fuel	0.3	225	0.3	90
Solid fuel	0.7	550	0.5 (subbituminous) 0.6 (bituminous)	100

a. ≥250 MBtu/hr heat input.
b. Expressed as NO_2.
c. Calculated at 3% excess O_2, dry basis.

Combustion modifications have their greatest effect on oil- and gas-fired boilers, where most of the NO_x is generated by the thermal mechanism, and less of an effect on coal-fired boilers, where most of the NO_x is formed from fuel nitrogen. With the current emphasis on coal as the fuel for all new, large boilers, combustion modifications may not be adequate in controlling the amount of NO_x emitted by stationary sources. To achieve the reduction of NO_x emissions which may become necessary in the future, FGT may be required.

PROJECT SCOPE

The primary purposes of this study are to compare the economics of selected types of NO_x FGT processes using a set of consistent premises and to provide current and more detailed information on the processes. This report consists of: (1) the design and economic premises, (2) a description of the processes and process combinations used in the economic evaluations, and (3) results of the economic evaluations and economic comparisons of systems.

Process Selection

The initial Phase I NO_x report (Technical Assessment of NO_x Removal Processes for Utility Application, EPA-600/7-77-127, EPRI AF-568, TVA Y-120), completed in late 1977, was a technical assessment of some 48 FGT processes currently listed in the literature. Each process was defined and evaluated by the following criteria: (1) process description, including a block flow diagram; (2) status of development; (3) background of the process developer;

(4) published economic data; (5) raw material, energy, and operation requirements; (6) technical considerations; (7) environmental considerations; (8) critical data gaps and poorly understood phenomena; and (9) overall technical advantages and disadvantages. In addition, each of the NO_x FGT processes was categorized as a wet- or a dry-treatment process and as a particular type of wet or dry process. These 48 processes described in the Phase I study are listed alphabetically in Table 4 as a wet or dry process and further as a particular wet or dry type. As seen from this table, a large proportion of the current FGT processes undergoing development belong to the dry selective catalytic reduction (SCR) type.

One purpose of the Phase I study was to select eight of the most promising FGT processes for preliminary economic analysis in this phase of the NO_x process study. These processes were selected on the basis of technical and economic considerations, development status, and the need for a representative sample.

The processes selected by these criteria are shown in Table 5. Four dry SCR processes were selected primarily on the basis of type of reactor (which is the only major piece of process equipment in dry NO_x-only systems) and because this type represents almost half of the processes under development. The Universal Oil Products, Inc., (UOP) Shell Flue Gas Treatment process has two variations: a simultaneous sulfur-oxide - nitrogen-oxide (SO_x-NO_x) process (SFGT-SN) and an NO_x-only process (SFGT-N), both of which were included. The two wet oxidation-absorption-reduction processes use different oxidants with potentially large cost differences. The Asahi Chemical process was selected as representative of wet absorption-reduction systems. The MON Alkali process, the eighth process recommended in the Phase I study, was deleted from this study because development work on the MON Alkali process was abandoned after publication of the Phase I report. The JGC Paranox process was considered for substitution as an eighth process. This process was investigated and correspondence ensued with UOP, the U.S. licensee. However, since this process technology is almost identical with the UOP SFGT-N process (the catalyst composition being the only difference), and the costs are claimed to be about the same as the UOP SFGT-N process, the JGC Paranox process was also omitted from this report.

To permit comparisons of NO_x-only and SO_x-NO_x FGT systems and to provide comparisons under a wider range of possible applications, a limestone scrubber flue gas desulfurization (FGD) unit is also included in the comparisons. This is used in conjunction with the NO_x-only FGT systems in comparisons with the SO_x-NO_x FGT systems. Also, the type and cost of particulate collection required will vary according to the type of FGT system. Therefore, to provide an equitable comparison between the NO_x FGT alternatives, particulate collection is also included in the comparisons.

Process Descriptions and Evaluations

The process descriptions in this report are amplified and updated versions of the material in the Phase I report (EPA-600/7-77-127; EPRI

TABLE 4. CURRENT FLUE GAS DENITRIFICATION PROCESSES

Dry processes	Wet processes
Selective catalytic reduction	Absorption-reduction
Asahi Glass[a]	Asahi Chemical
Eneron	Chisso Engineering
Exxon	Kureha
Hitachi, Ltd.	Mitsui Engineering and Shipbuilding
Hitachi Zosen	Pittsburgh Environmental and Energy Systems
JGC Paranox	Oxidation-absorption-reduction
Kobe Steel	Chiyoda Thoroughbred 102
Kurabo Knorca	Ishikawajima-Harima Heavy Industries
Kureha	Mitsubishi Heavy Industries
Mitsubishi Heavy Industries	Moretana Calcium
Mitsubishi Kakoki Kaisha	Moretana Sodium
Mitsubishi Petrochemical	Osaka Soda[a]
Mitsui Engineering and Shipbuilding	Shirogene[a]
Mitsui Toatsu	Absorption-oxidation
Nippon Kokan[a]	Hodogaya[a]
Sumitomo Chemical	Kobe Steel
Sumitomo Heavy Industries[b]	MON Alkali Permanganate
Takeda	Nissan Engineering
Ube	Oxidation-absorption
Unitika[b]	Kawasaki Heavy Industries
Universal Oil Products-Shell[b]	Tokyo Electric-Mitsubishi HI
Nonselective catalytic reduction	Ube
The Ralph M. Parsons	
Selective noncatalytic reduction	
Exxon Thermal Denox	
Adsorption	
Foster Wheeler-Bergbau Forschung	
Radiation	
Ebara-JAERI	
Uncertain	
Princeton Chemical Research[a]	

a. Essentially no information available.

b. These companies have two dry processes: (1) NO_x-only and (2) SO_x-NO_x.

TABLE 5. PROCESSES SELECTED FOR FURTHER STUDY IN PHASE II

Process	Type of process (classification)
UOP Shell Flue Gas Treatment (SFGT-N)	Dry NO_x only (SCR)
Hitachi Zosen	Dry NO_x only (SCR)
Kurabo Knorca	Dry NO_x only (SCR)
UOP Shell Flue Gas Treatment (SFGT-SN)	Dry simultaneous SO_x-NO_x (SCR)
Moretana Calcium	Wet simultaneous SO_x-NO_x (oxidation-absorption-reduction)
Ishikawajima-Harima Heavy Industries (IHI)	Wet simultaneous SO_x-NO_x (oxidation-absorption-reduction)
Asahi Chemical	Wet simultaneous SO_x-NO_x (absorption-reduction)

AF-568; TVA Y-120). In addition, flow diagrams and material balances are included. The design and economic premises are similar to those used in previous Tennessee Valley Authority (TVA) FGD studies, with the inclusion of additional premises for NO_x FGT. These premises define design and operating conditions for a representative power unit, permitting comparison of the systems evaluated on an equal basis.

The economic evaluation consists of comparisons of estimated total capital investment and average annual revenue requirements for each process and process combination. Estimates were prepared and comparisons made for the following systems: (1) wet versus dry SO_x-NO_x FGT systems, (2) dry NO_x-only systems with limestone FGD (SCR-FGD) versus wet SO_x-NO_x FGT systems, (3) SCR-FGD versus dry SO_x-NO_x FGT systems, (4) wet SO_x-NO_x FGT systems using oxidation-absorption-reduction steps versus those using absorption-reduction steps, and (5) moving-bed reactor versus parallel flow, fixed-bed reactors for dry NO_x FGT.

DESIGN AND ECONOMIC PREMISES

An accurate comparison of the various types of NO_x FGT processes can be made only if a specific set of design and economic premises is established as a basis for the design and economic calculations. In addition, an analysis of various NO_x FGT processes is complicated because the dry SCR processes are NO_x-only FGT systems. Therefore, the NO_x-only systems have been combined with a conventional limestone FGD system so that all the alternatives are compared on an equal basis (i.e., minimum 90% removal of both SO_x and NO_x). In addition, particulate collection, the type and cost of which vary between the different NO_x FGT processes, is included for an equitable, overall comparison of SO_x-NO_x-particulate matter (PM) control systems. The pertinent design and economic premises concerning the operation of the power plant, the FGD system, the various electrostatic precipitators (ESP), and the NO_x FGT systems are discussed in this section.

DESIGN PREMISES

The basic design premises for power plant stack gas emission control studies were established by TVA and EPA for work conducted prior to this study. These premises are similar to those used in other technical and economic evaluations of power plant emission control technology. After consultation between TVA, EPA, and Electric Power Research Institute (EPRI), these premises, with appropriate additions and minor adjustments (primarily dealing with NO_x control), were also used for this evaluation. Using these general design premises, a set of base case design premises was established.

Power Plant

Plant Size and Fuel--

The only power plant evaluated in this study was a new, 500-MW coal-fired boiler. A midwestern location was assumed as a basis for the power unit because of the concentration of power stations in that area and the proximity of this area to the major coal fields. The fuel for the plant was a coal having a heating value of 10,500 Btu/lb and containing 3.5% sulfur and 16% ash. The coal composition and the input coal requirements (based on a heat rate of 9,000 Btu/kWh) for the 500-MW boiler are listed in Table 6.

TABLE 6. BASE CASE COAL COMPOSITION

AND INPUT FLOW RATE

(500-MW new unit, 9,000 Btu/kWh heat rate)

Component	Wt %, as fired	Lb/hr
C	57.56	246,800
H_2	4.14	17,700
N_2	1.29	5,500
O_2	7.00	30,000
S	3.12	13,400
Cl	0.15	600
Ash	16.00	68,600
H_2O	10.74	46,000
Total	100.00	428,600

Power Unit Life, Operating Time, and Capacity--

Based on guidelines suggested by the then Federal Power Commission (FPC) (7), a new, coal-fired power plant is estimated to have a 30-year life. However, the operating capacity for these plants varies widely over their lifetime. Power unit operating schedules that reflect TVA experience were used in this study. These are shown in Table 7.

TABLE 7. ASSUMED POWER PLANT CAPACITY SCHEDULE

Operating yr	Capacity factor, % (nameplate rating)	Annual operating time, hr
1-10	80	7,000
11-15	57	5,000
16-20	40	3,500
21-30	17	1,500
Average for 30-yr life	48.5	4,250

Flue Gas Composition--

Because dry SCR processes operate at higher temperatures (570-780°F) than wet processes, most process developers recommend that the flue gas be removed at the outlet of the boiler economizer for passage through the FGT system. Flue gas for the wet FGT processes is removed at the outlet of the boiler air heater. Because of air leakage in the boiler air heater--assumed to be 13% of the stoichiometric combustion air requirements--the flue gas composition at this position is different than the composition of the flue gas at the economizer outlet. The flue gas composition at the economizer outlet is shown in Table 8 and at the boiler air heater outlet in Table 9.

TABLE 8. FLUE GAS COMPOSITION AND FLOW RATE AT THE ECONOMIZER OUTLET

(500-MW new unit, coal-fired, 9,000 Btu/kWh,
3.5% S, 10,500 Btu/lb HHV as fired,
2,146,000 aft^3/min at 705°F)

Component	Vol, %	Lb/hr
N_2	73.39	3,113,000
O_2	3.23	156,500
CO_2	13.56	904,200
SO_2	0.26	25,130
SO_3	0.003	317
NO_x	0.064	3,009
HCl	0.012	661
H_2O	9.48	258,600
Total	100.00	4,461,000 (approx)

TABLE 9. FLUE GAS COMPOSITION AND FLOW RATE AT AIR HEATER OUTLET

(500-MW new unit, coal-fired, 9,000 Btu/kWh,
3.5% S, 10,500 Btu/lb HHV as fired,
1,543,000 aft^3/min at 300°F)

Component	Vol, %	Lb/hr
N_2	73.76	3,450,000
O_2	4.83	258,200
CO_2	12.31	904,200
SO_2	0.24	25,130
SO_3	0.0024	317
NO_x	0.060	3,009
HCl	0.01	661
H_2O	8.79	264,500
Total	100.00	4,906,000 (approx)

The flue gas compositions are calculated on the basis that 95% of the sulfur in the coal is emitted in the flue gas as SO_x [99% sulfur dioxide (SO_2) and 1% sulfur trioxide (SO_3)] and 80% of the ash is emitted in the flue gas as particulates. These values represent TVA operating experience with frontal-fired coal-burning units.

The NO_x concentration of the flue gas at both positions is 600 ppm by volume and consists of about 95% nitric oxide (NO) and 5% nitrogen dioxide (NO_2). This concentration and composition is believed representative of the range of concentrations likely to occur in particular applications.

Emission Standards--

Although the pre-1979 NSPS for new, coal-fired utility boilers were in lb/MBtu heat input the 1977 amendments to the Clean Air Act of 1970 required establishment of NSPS that specify a percent reduction as well as a maximum emission rate with respect to SO_2, which was accomplished in the 1979 NSPS. Therefore, each of the FGT process packages in this study includes all the equipment required to achieve particulate, SO_x, and NO_x removal efficiencies of 99.5%, 90%, and 90% respectively. Since the wet FGT processes are capable of removing more than 90% of the inlet SO_x, a limestone FGD system capable of 90% SO_x removal was coupled with the NO_x-only FGT systems to form comparable overall systems. An ESP system was coupled with certain FGT systems to form a comparable overall system with the wet FGT processes using wet scrubbers for particulate and chloride removal.

FGT Process Premises

For this study the ducts from the boiler air heaters are assumed to exhaust to a common plenum connecting the scrubbing trains. Separate ducts from the plenum to each scrubbing train are equipped with dampers for individual scrubber shutoff during periods of maintenance or power plant turndown.

With the exception of the Asahi process where flue gas from an oil-fired heater is available, and the UOP SFGT-SN process which is a dry SO_x-NO_x system upstream of the air heater-stack gas reheat to 175°F for plume buoyancy is provided by indirect steam reheat of the cleaned gas. For reheat calculations the amount of entrained water in the flue gas to the reheater is assumed to be 0.1 weight percent of the wet-gas mass flow rate in each processing system.

Separate induced draft (ID) fans are not included for the dry FGT systems. Rather, a larger boiler ID fan is used to compensate for the higher pressure drop of the combined boiler-FGT system and the increased costs are assigned to the FGT process. Separate booster fans and ductwork for the SO_x-NO_x systems and the limestone FGD unit are included because of the larger pressure drops involved with these systems.

For systems requiring disposal of sludge or solid waste other than fly ash, a disposal pond or landfill located 1 mile from the plant site is provided. These are designed to minimize land and construction costs and are sized for a 30-year life at the waste flow rates of the process.

In addition, the design premises assume proven technology, not first-of-a-kind installations. Other than spares for pumps no redundancy features are included. In actual operation, additional spares or redundancy may be required.

There are certain items which are excluded in this report. The investment and revenue requirement for the fly ash pond, which should be similar for each process, are omitted. Also excluded are the costs associated with additional soot blowing which may be necessary for the air heater with the dry FGT processes.

Costs for the ash-handling system to move the fly ash to final disposal and soot blowing of the dry FGT process reactors are included. Also included are any incremental cost differences required to recover claimed heat credits by the dry processes, for example, larger air heaters.

Two reactor trains were selected for the dry processes, except for the Kurabo Knorca system in which four trains were recommended, to provide a comparative basis.

Asahi Process--

The Asahi process is an example of the wet, simultaneous SO_x-NO_x, absorption-reduction type FGT system. The relatively low mass transfer coefficients associated with the absorption of NO by aqueous solutions make large absorbers necessary. Thus the initial sections of the Asahi process, prescrubbing and absorption, consist of four separate trains for a 500-MW coal-fired boiler. The prescrubbers are modified venturi scrubbers and the absorbers are packed towers. The design conditions and removal efficiencies in the prescrubbers and absorbers are shown in Table 10.

Although there are four trains of prescrubbers and absorbers, most of the regeneration section is a single train. To avoid potential forced boiler outages due to the regeneration section going off-line for routine maintenance or repairs, excess surge capacity (usually 8 hours) is included at strategic locations throughout the regeneration system. The process design includes all equipment needed to dispose of the byproduct gypsum in a landfill. Sodium sulfate (Na_2SO_4) is assumed to be sold at no profit.

Ishikawajima-Harima Heavy Industries (IHI) Process--

The IHI process is a wet, simultaneous SO_x-NO_x, oxidation-absorption-reduction type FGT system. In this type of FGT process, a gas-phase oxidant is used to convert NO to NO_2. Since NO_2 is absorbed much more readily than NO, the developers of the IHI process believe that only a single train of processing equipment for a 500-MW coal-fired boiler is required. However, to prevent boiler outages due to shutdown of the FGT system, additional surge volume (usually 8 hours) has been included in each section of the regeneration system.

TABLE 10. DESIGN CONDITIONS AND REMOVAL EFFICIENCIES FOR THE PRESCRUBBER AND THE ABSORBER IN THE ASAHI SO_x-NO_x PROCESS

Item	Prescrubber	Absorber
Type	Modified venturi	Tellerette packed-bed
Operating conditions		
Pressure drop, in. of H_2O	7.9	7.9
L/G, gal/kaft^3	6.2	62.2
pH	1.0	6.0-6.5
Removal efficiency, %[a]		
SO_2	10.0	99.0
SO_3	25.0	86.0
NO_x	-	90.0
HCl	99.0	90.0
Ash	99.0	90.0

a. [(Inlet-outlet)/inlet]100.

The single prescrubber and absorber have the design conditions and removal efficiencies shown in Table 11. Essentially all of the absorbed SO_x is converted to gypsum. From initial development work it appears that approximately 80% of the absorbed NO_x is converted to complex nitrogen-sulfur compounds, 15% is converted to molecular nitrogen, and the remaining 5% remains as nitrite-nitrate salts. Therefore these conversion rates were assumed for this study.

TABLE 11. DESIGN CONDITIONS AND REMOVAL EFFICIENCIES FOR THE PRESCRUBBER AND THE ABSORBER IN THE IHI SO_x-NO_x PROCESS

Item	Prescrubber	Absorber
Type	Modified venturi	Spray tower
Operating conditions		
Pressure drop, in. of H_2O	12.0	4.0
L/G, gal/kaft^3	NA[a]	NA[a]
pH	1.0-2.0	4.0-5.0
Removal efficiency, %[b]		
SO_2	2.0	90.0
SO_3	-	-
NO_x	-	90.0
HCl	95.0	0.0
Ash	99.5	0.0

a. Not available, proprietary information.
b. [(Inlet-outlet)/inlet]100.

The ozone for the IHI process is generated by corona discharge which produces 1.0 weight percent ozone in the airstream from the generators. The ozone generation system capacity is based on a required ozone to NO_x mol ratio of 0.85:1.0 (assuming 5% of the NO_x is already NO_2) in the flue gas. The process design includes a landfill site for the disposal of the byproduct gypsum and mixed salts.

Moretana Calcium Process--

The Moretana Calcium process is also a wet, simultaneous SO_x-NO_x, oxidation-absorption-reduction process. Although the developers of the IHI process believe only one train is needed for a 500-MW coal-fired boiler, Sumitomo Metal, developer of the Moretana Calcium process, recommends two trains of equipment. Even though the process developers recommend two trains, only one train of limestone preparation equipment is used and sufficient surge storage is designed into the limestone slurry tank.

The Moretana tower absorber is designed for the operating conditions and removal efficiencies given in Table 12. Most of the SO_x absorbed is converted to a mixed sulfite-sulfate sludge similar to that produced in a conventional limestone FGD system. The NO_x, on the other hand, is oxidized to both NO_2 and nitric acid (HNO_3) and after absorption is converted to molecular nitrogen and calcium nitrate [$Ca(NO_3)_2$] respectively. This conversion of NO_x to nitrogen and $Ca(NO_3)_2$ is assumed to be equimolar, with no other nitrogen byproducts formed.

TABLE 12. DESIGN CONDITIONS AND REMOVAL EFFICIENCIES FOR THE PRESCRUBBER AND THE ABSORBER IN THE MORETANA CALCIUM SO_x-NO_x PROCESS

Item	Prescrubber	Absorber
Type	Moretana tower	Moretana tower
Operating conditions		
Pressure drop, in. of H_2O	3.9	8.3
L/G, gal/kaft^3	30.8	87.1
pH	1.0	5.0
Removal efficiencies, %[a]		
SO_2	-	95.0
SO_3	50.0	50.0
NO_x	-	90.0
HCl	95.0	100.0
Ash[b]	90.0	80.0
ClO_2	-	100.0
Cl_2	-	100.0
HCl	-	100.0
HNO_3	-	100.0

a. [(Inlet-outlet)/inlet]100.

b. Moretana Calcium process is equipped with a 94% efficient ESP prior to prescrubber.

The design mol ratio of chlorine dioxide (ClO_2) to NO_x is 0.55:1.0 for 90% NO_x removal. Equipment is included in each ClO_2 generating unit for adequate surge storage of ClO_2. The process design also includes a sludge pond for the disposal of sulfite-sulfate sludge.

Hitachi Zosen Process--

The Hitachi Zosen process is a dry, NO_x-only, SCR system with two processing trains and using a fixed-bed reactor and corrugated-shaped catalyst. The ammonia (NH_3) is diluted with air to 5% by volume for easier flow control, enhanced mixing, and because the limits of inflammability for NH_3 in air are 15.5 to 27.0%. The design conditions and removal efficiency in the reactor are shown in Table 13. The overall process design for the Hitachi Zosen system includes a limestone FGD system and a high-efficiency cold ESP.

TABLE 13. DESIGN CONDITIONS AND REMOVAL EFFICIENCIES FOR THE CATALYTIC REACTOR IN THE HITACHI ZOSEN NO_x-ONLY PROCESS

Item	Reactor
Type	Fixed-bed (corrugated)
Operating conditions	
Pressure drop, in. of H_2O	2.0-3.0
Space velocity, $ft^3/hr/ft^2$	5,000-10,000
NH_3:NO_x mol ratio	(1.0-1.2):1.0
Temperature, °C	300-400
Catalyst material	Proprietary
NO_x removal efficiency, %[a]	90.0

a. [(Inlet-outlet)/inlet]100.

Kurabo Knorca Process--

The Kurabo Knorca process is also a dry, NO_x-only, SCR process. However, in contrast to the other NO_x-only FGT processes, the Knorca process has four reactor trains in a 500-MW coal-fired power plant. Also, a moving-bed reactor is used. The NH_3 injected into the flue gas is not diluted, but rather, static gas mixers are included in the flue gas ducts to ensure adequate mixing of the gases. The design conditions and removal efficiency in the reactor are shown in Table 14. The overall design system includes a limestone FGD system and a hot ESP.

UOP SFGT-N Process--

The UOP SFGT-N dry, NO_x-only, SCR process consists of two processing trains in a 500-MW coal-fired application with a parallel passage, fixed-bed reactor. The design conditions and removal efficiencies in the catalytic reactors are shown in Table 15. In order to ensure good mixing and flow

TABLE 14. DESIGN CONDITIONS AND REMOVAL EFFICIENCIES FOR THE CATALYTIC REACTOR IN THE KNORCA NO_x-ONLY PROCESS

Item	Reactor
Type	Moving-bed
Operating conditions	
Pressure drop, in. of H_2O	4
Space velocity, $ft^3/hr/ft^2$	7,500
NH_3:NO_x mol ratio	0.9:1.0
Temperature, °C	400
Catalyst material	Fe-oxide on TiO_2
NO_x removal efficiency, %[a]	90.0

a. [(Inlet-outlet)/inlet]100.

TABLE 15. DESIGN CONDITIONS AND REMOVAL EFFICIENCIES FOR THE CATALYTIC REACTOR IN THE UOP SFGT-N, NO_x-ONLY PROCESS

Item	Reactor
Type	Parallel passage
Operating conditions	
Pressure drop, in. of H_2O	5.0-6.0
Space velocity, $ft^3/hr/ft^2$	7,500
NH_3:NO_x mol ratio	1.1:1.0
Temperature, °C	400
Catalyst material	CuO and $CuSO_4$ on alumina
NO_x removal efficiency, %[a]	90.0

a. [(Inlet-outlet)/inlet]100.

control and because of inflammability limits, the NH_3 injected is first diluted with air to 1.0 weight percent (5% by volume) NH_3. The overall process design for the UOP SFGT-N system includes a limestone FGD system and a high-efficiency cold ESP.

UOP SFGT-SN Process--

The UOP SFGT-SN process is the only dry, simultaneous SO_x-NO_x, SCR process included in this study and uses a parallel passage, fixed-bed reactor. With the exception of the eight reactors, one train of processing equipment is used in a 500-MW coal-fired application. The design conditions and removal efficiencies in the reactors are shown in Table 16.

TABLE 16. DESIGN CONDITIONS AND REMOVAL EFFICIENCIES FOR THE REACTOR IN THE UOP SFGT-SN, SO_x-NO_x PROCESS

Item	Reactor
Type	Parallel passage
Operating conditions	
Pressure drop, in. of H_2O	15.0-16.0
Space velocity, $ft^3/hr/ft^2$	7,500
NH_3:NO_x mol ratio	1.1:1.0
Temperature, °C	400
SO_2 sorbent material	CuO on alumina
NO_x catalyst material	$CuSO_4$ on alumina
Removal efficiency, %[a]	
SO_2	90
SO_3	70
NO_x	90

a. [(Inlet-outlet)/inlet]100.

The process design includes the equipment required to convert the absorbed SO_2 to 98% H_2SO_4. Normally the ESP would be considered as part of the boiler cost; however, the cost of the ESP is included in this study in order to compare the process with wet FGT systems, which have a prescrubber for particulate removal.

Limestone FGD--

The limestone FGD system with four trains of absorbers is designed for 90% removal of SO_2 only; it includes neither an ESP nor a prescrubber. Each of these scrubbers is a Turbulent Contact Absorber (TCA) type with a presaturator for cooling and humidifying the flue gas. The design removal efficiencies for both the presaturator and the absorber are given in Table 17. Each scrubber is equipped with a chevron-type entrainment separator at the scrubber outlet. Stack gas reheat to 175°F for plume buoyancy is provided by indirect steam reheat of the cleaned gas.

Waste sludge is disposed of by ponding as a 15 weight percent solids slurry. The settled sludge contains 60% free water; the excess water is returned to the scrubber system.

ESP--

Because of the various ESP locations and PM removal efficiencies required with the different NO_x FGT processes, four different ESP cases are included in this study. With three of the dry NO_x FGT processes, i.e., Hitachi Zosen, UOP SFGT-N, and UOP SFGT-SN, a cold ESP is used which is located just downstream of the air heaters. Four trains of ESP are used. With the UOP SFGT-N and Hitachi Zosen processes, the ESP units are sized and costs assigned on the basis of a 99.5% PM removal efficiency and a specific collecting area of 300 ft^2 of collecting electrode per kft^3/min of flue gas. With the UOP SFGT-SN process, since some SO_3 is also removed, removing fly ash is more difficult. For 99.5% PM removal efficiency, the units are sized and costs assigned on the basis of a specific collecting area of 460 ft^2 of collecting electrode per kft^3/min of flue gas. With the Kurabo Knorca process, the remaining dry NO_x FGT process, a hot ESP upstream of the NO_x reduction reactor is used. Four ESP units are used. These units are sized and costs assigned on the basis of a 98% PM removal efficiency and specific collecting area of 240 ft^2 of collecting electrode per kft^3/min of flue gas. The remaining PM removal is achieved by the NO_x reduction reactor for a total of 99.5% removal efficiency. The only wet NO_x FGT process using an ESP is the Moretana Calcium process. A PM removal efficiency of 99.5% is achieved by using a 94% efficient cold ESP followed by a prescrubber. A single ESP is included. The ESP is sized and costs are assigned on the basis of a specific collecting area of 150 ft^2 of collecting electrode per kft^3/min of flue gas.

TABLE 17. DESIGN OPERATING CONDITIONS AND REMOVAL EFFICIENCIES FOR THE PRESATURATOR AND THE ABSORBER IN THE LIMESTONE FGD SYSTEM

Item	Presaturator	Absorber
Operating conditions		
Pressure drop, in. of H_2O	-[a]	8.0
L/G, gal/$kaft^3$	4.0	50.0
Removal efficiency, %[b]		
SO_2	5	90
SO_3	50	50
HCl	100	-
NO_x	0	0

a. Included in the pressure drop for the absorber.
b. Defined as [(inlet-outlet)/inlet]100.

Raw Materials--

All raw materials used in the eight NO_x FGT systems are assumed to be received by either rail or truck. Thirty-day storage facilities are provided for them. Feed bins and intermediate process tanks are designed for an 8-hour storage capacity.

Listed below are the primary raw materials and their typical properties.

1. Limestone (Asahi, IHI, Moretana Calcium, and limestone FGD)
   Analysis: 90% calcium carbonate

2. Na_4·EDTA (Asahi)
   Analysis: 39% aqueous solution of Na_4·EDTA

3. Soda ash (Asahi)
   Analysis: 98% Na_2CO_3

4. Sulfuric acid (Asahi, IHI)
   Analysis: 98% H_2SO_4

5. Caustic soda (IHI)
   Analysis: 50% NaOH

6. Sodium chloride (IHI)
   Analysis: 97% NaCl

7. Slaked lime (Moretana Calcium)
   Analysis: 93% $Ca(OH)_2$

8. Ammonium sulfate (Moretana Calcium)
   Analysis: 99% $(NH_4)_2SO_4$

9. Hydrogen chloride (Moretana Calcium)
   Analysis: 35% HCl

10. Ammonia (Hitachi Zosen, Kurabo, UOP)
    Analysis: Guaranteed 99.5% NH_3, 82.2% N

11. Naphtha (UOP SFGT-SN)
    Analysis: 133,000 Btu/gal

ECONOMIC PREMISES

The following criteria provide a common basis for the economic evaluation of each of the eight alternative NO_x FGT processes and make the results directly comparable.

Capital Investment

Capital investment estimates are based on a midwestern plant location and represent projects beginning in mid-1977 and ending in mid-1980, with an average cost basis for scaling of mid-1979. Direct investments are calculated using the average annual *Chemical Engineering* cost indices for the period up to 1975 and the recent TVA projections for the period 1976-1981 shown in Table 18. Actual equipment cost estimates are based on current (1978) cost information obtained from engineering-contracting, processing, and equipment companies.

For the dry FGT systems, only the incremental investment and operating cost for the larger ID fans required is charged to the FGT system. This incremental cost is estimated as the difference between the larger ID fan required for the combined boiler-FGT system pressure drop and the ID fan which would be needed if no NO_x removal facilities were installed. This option is usually less expensive than the addition of a separate fan to compensate for the pressure drop through the NO_x reactor.

Costs related to equipment, material, and construction-labor shortages with accompanying overtime pay incentive and costs for the generation facilities for electricity used by the NO_x removal processes and FGD unit are not included.

TABLE 18. COST INDICES AND PROJECTIONS

Year	1970	1971	1972	1973	1974	1975	1976[a]	1977[a]	1978[a]	1979[a]	1980[a]	1981[a]
Plant	125.7	132.3	137.2	144.1	165.4	182.4	197.9	214.7	232.9	251.5	271.6	293.3
Material[b]	123.8	130.4	135.4	141.9	171.2	194.7	210.3	227.1	245.3	264.9	286.1	309.0
Labor[c]	137.4	146.2	152.2	157.9	163.3	168.6	183.8	200.3	218.3	237.9	259.3	282.6

a. Projections.
b. Same as index in *Chemical Engineering* for "equipment, machinery, supports."
c. Same as index in *Chemical Engineering* for "construction labor."

Direct Investment--

The purchase cost of the process equipment and the cost of materials and labor for the installation of this equipment are included. Also, the cost of piping and insulation, ductwork, excavation, site preparation, foundation, structural, roads and railroads, electrical, instrumentation, buildings, painting, services, and pond construction required for each unit area are estimated. The costs for the NO_x FGT processes and ESP's are estimated based on the flow diagram, equipment lists, and TVA experience. The calculation of the direct investment for the limestone FGD system, on the other hand, is based on detailed engineering design.

Services, utilities, and miscellaneous costs are estimated at 6% of the total direct investment excluding pond construction cost. This expense covers allocated costs for the use of such power plant facilities as maintenance shops, stores, communications, security, and offices. Parking lots, walkways, landscaping, fencing, and vehicles are also included in the service facility estimate.

Indirect Investment--

The indirect investment includes costs for engineering design and supervision, architect and engineering contractor expenses, construction expenses, contractor fees, and contingency. The engineering design and supervision and contingency factors are based on proven design, not first-of-a-kind installation.

Engineering design and supervision--This indirect investment factor was estimated using a technique that correlates the number of major pieces of process equipment with drafting room man-hour and engineering design costs. Pond engineering and design costs were determined by a separate procedure based on pond construction expense. In a similar manner, the engineering design and supervision costs for landfill disposal cases are based on major items excluding earthmoving equipment. The sum of these costs, where applicable, appears in the indirect investment display as engineering design and supervision for each process.

Architect and engineering (A&E) contractor expenses--This cost is based directly upon the engineering design and supervision costs. A&E expense is assumed to be 25% of the portion of engineering design and supervision costs associated with major equipment. For cases involving disposal ponds, 10% of the engineering design and construction expenses associated with pond construction is estimated as the additional A&E expense. For the landfill systems, 25% of the total engineering design and supervision expense is assumed to be the A&E expense.

Construction expense--Construction expense is expressed as a function of the direct investment and was estimated by the following equation.

$$\text{Construction expense} = 0.25\ (x)^{0.83} + 0.13\ (y)^{0.83}$$

where x = direct investment, excluding pond or landfill investment cost, in M$

y = direct pond cost in M$

Construction facilities (which include costs for mobile equipment, temporary lighting, construction roads, raw water supply, safety and sanitary facilities, and other similar expenses incurred during construction) are considered a part of construction expenses.

Contractor fees--The relationship between contractor fees and total direct investment used to estimate the cost of contractor fees was:

$$\text{Contractor fees} = 0.096\ (a)^{0.76}$$

where a = direct investment in M$

Contingency--The contingency is assumed to be 20% of the sum of total direct investment, engineering design and supervision costs, A&E expenses, construction expenses, and contractor fees.

Other Capital Charges--

The sum of the total direct and total indirect investment is total fixed investment. Other capital charges which must be added to this total fixed investment to obtain the total capital investment are: (1) allowance for startup and modifications, (2) interest during construction, (3) land, (4) working capital, and (5) royalty fees.

Allowance for startup and modifications--This expense is estimated to be 10% of the total fixed investment excluding pond construction.

Interest during construction--This item was estimated to be 12% of the total fixed investment. This percentage is calculated as the simple interest which would be accumulated at a 10%/yr rate assuming a debt-equity ratio of 60:40 on the incremental capital investment and a 3-year project expenditure schedule as shown in Table 19.

Land--The cost of land is estimated at $3500/acre.

Working capital--Working capital consists of: (1) money invested in raw materials, supplies, and finished products carried in stock and semi-finished products in the process of being manufactured; (2) accounts receivable; (3) cash retained for payment of operating expenses, such as salaries, wages, and raw material purchases; (4) accounts payable; and (5) taxes payable (14). For these premises, working capital is defined as the equivalent cost of 3 weeks of raw material costs, 7 weeks of direct costs, and 7 weeks of overhead costs.

Royalty fees--The royalty fees were added based on information supplied by the vendor. If there were no data, a $0.3M fee was assumed. For both UOP processes, the royalty fee was prorated for inclusion in the direct investment.

TABLE 19. PROJECT EXPENDITURE SCHEDULE

	Year			
	1	2	3	Total
Fraction of total expenditure as borrowed funds	0.15	0.30	0.15	0.60
Simple interest at 10%/yr as % of total expenditure				
Year 1 debt	1.5	1.5	1.5	4.5
Year 2 debt	-	3.0	3.0	6.0
Year 3 debt	-	-	1.5	1.5
Accumulated interest as % of total expenditure	1.5	4.5	6.0	12.0

Revenue Requirements

Annual revenue requirement calculations are based on 7000 hours of operation per year using mid-1980 operating costs and average capital charges. Process operation schedules are assumed to be the same as the power plant operating profiles and remaining life assumptions given in the power plant design premises.

Direct Costs--

Direct operating costs include raw material, labor, utility, and maintenance costs. The 1980 projected unit costs for the raw materials used in the various NO_x FGT processes are given in Table 20 as delivered costs to a midwestern location. Because of the proprietary nature of the catalyst used in some of these FGT processes, the costs for these catalyst are listed separately and are given only as annual quantities in Table 20. The catalyst cost is based on vendor quotes and a guaranteed life of 1 year. The projected labor costs assumed for 1980 are shown in Table 21.

Projected 1980 utility costs are shown in Table 22. Unit costs for electricity and steam generated by the power plant are based on actual production cost including labor, fuel, depreciation, taxes, and rate base return on investment. The electricity rates are based on purchase from an independent source with full capital recovery provided and are adjusted for the quantity used. Process water rates vary with the quantity of water required. If a heat credit is applied to a process, $2/MBtu is used as a basis for the credit.

Maintenance costs are estimated as a percentage of the direct investment, excluding pond construction. The maintenance factors vary according to operating characteristics of the process and are estimated based on either actual operating experience or maintenance needs in similar process areas. The estimated maintenance factors for each process are shown in

TABLE 20. PROJECTED 1980 COSTS FOR RAW MATERIALS

Raw material	Cost, \$/ton
Limestone	7.00
Na_2CO_3	103.00
$Na_4 \cdot EDTA$	743.00
H_2SO_4	54.00
K_2SO_4	144.00
Copperas	86.50
Slaked lime	60.00
$CuCl_2$	2,105.00
HCl	81.00
$(NH_4)_2SO_4$	82.00
Powdered limestone	32.00
NaCl	25.80
NaOH	187.00
NH_3	150.00
Naphtha	285.60[a]
Catalyst	**Cost, \$/yr**
Moretana Calcium	4,116,200
Hitachi Zosen	6,880,000
Kurabo	4,316,000
UOP SFGT-SN	6,648,000
UOP SFGT-N	2,449,000

a. Based on \$0.50/gal with 42 gal/bbl and 13.6 bbl/ton.

TABLE 21. PROJECTED 1980 UNIT COST FOR LABOR

Type	Unit cost, \$/man-hr
Plant operation and supervision	12.50
Analyses	17.00
Landfill equipment and truck operators	17.00

Table 23. Pond maintenance is calculated separately as 3% of the pond construction cost. The maintenance charge for the ozone system in the IHI process was assumed to be 4% of ozone plant investment. Maintenance for the landfill section was also assumed to be 4% of the total investment for the landfill.

TABLE 22. PROJECTED 1980 UNIT COST FOR UTILITIES

Utility	Cost, $
Steam	2.00/MBtu
Electricity	0.029/kWh
Process water (UOP)	0.06/kgal
Process water	0.12/kgal
Naphtha	0.50/gal

TABLE 23. ESTIMATED ANNUAL MAINTENANCE FACTOR

Process	Percentage of direct investment
Asahi	7
Hitachi Zosen	4
IHI	7[a]
Kurabo	4
Moretana Calcium	7[b]
UOP SFGT-SN	4
UOP SFGT-N	4
Limestone FGD	8[b]

a. Maintenance of ozone section of IHI is 4% of investment for ozone manufacturing plant.

b. Excluding pond construction. Pond maintenance is estimated as 3% of pond construction cost and maintenance for landfill is assumed to be 4% of investment.

Other direct costs such as land preparation cost (i.e., the annual expense for clearing, excavating, draining, and reclaiming the disposal site) are based on the annual quantity of land used, and are assumed to be $1700/acre. The operating cost of trucks and landfill equipment (fuel and maintenance) for a 1-mile travel distance is estimated at $0.06/ton of wet sludge. A fuel and maintenance cost of $0.16/ton of wet sludge is assumed for earthmoving equipment used in the landfill operation (2).

Indirect Costs--

Indirect costs include capital charges, projected to 1980, and overheads. Following power industry practice, regulated company economics are used for calculating the capital charges. A breakdown of the capital charges is given in Table 24. The depreciation rate is straight line based on the life of the power plant.

TABLE 24. ANNUAL CAPITAL CHARGES FOR POWER INDUSTRY FINANCING

	Percentage of total depreciable capital investment
Years remaining life	30
Depreciation-straight line (based on years remaining life of power unit)	3.3
Interim replacements (equipment having less than 30-yr life)	0.7
Insurance and property taxes	2.0
Total rate applied to original investment	6.0
	Percentage of unrecovered capital investment[a]
Cost of capital (capital structure assumed to be 60% debt and 40% equity)	
Bonds at 10% interest	6.0
Equity[b] at 14% return to stockholder	5.6
Income taxes (Federal and State)[c]	5.6
Total rate applied to depreciation base	17.2[d]

a. Original investment yet to be recovered or "written off."
b. Contains retained earnings and dividends.
c. Since income taxes are approximately 50% of gross return, the amount of taxes is the same as the return on equity.
d. Applied on an average basis, the total annual percentage of original fixed investment for new (30 yr) plants would be 6.0% + 1/2 (17.2%) = 14.6%.

In estimating the regulated capital charges associated with FGT, the conventional method of considering the overall life of the power plant is used. FPC recognizes the conclusion of the National Power Survey that a 30-year service life is reasonable for steam-electric plants. Because some life spans are less than 30 years, however, FPC has designated interim replacements as an allowance factor to be used in estimating annual revenue requirements. Use of this allowance following FPC-recommended practice provides for financing the cost of replacing such short-lived units. An average allowance of about 0.35% of the total investment is normally provided. However, to provide for the unknown life span of SO_x-NO_x control facilities, a larger allowance factor of 0.70% is used for new units. An insurance allowance of 0.5% is also included in the capital charges based on FPC practice. Property taxes are estimated at 1.5% of the total depreciable capital investment.

Debt-equity ratio is another component of capital charges for which variations of ratios may be expected. FPC data indicate that the long-term debt for privately owned electric utilities varied only slightly from 51.5 to 54.8% of total capitalization during the period 1965-1973. However, recent economic trends have changed the incremental debt-equity ratio because utilities are more dependent on bonds and bank loans for project funding. For this study, the capital structure is assumed to be 60% debt and 40% equity. The interest rate for bonds is assumed to be 10% and the return to stockholders on equity 14%. Costs of capital and income tax charges are applied to the uncovered portion of capital investment. Income taxes and return on equity are each about 50% of gross return. Since return on equity is 5.6% of total capital investment, income taxes are also 5.6% of total capital investment (see Table 24). Since most regulatory commissions base the annual permissible return on investment on the remaining depreciation base (that portion of the original investment yet to be recovered or "written off"), a portion of the annual capital charge included in the lifetime operating costs declines uniformly over the life of the power plant.

Plant, administrative, and marketing overheads vary from company to company. Based on the various methods used in industry and illustrated in a variety of cost estimating sources, the following method of estimating overheads is used. Plant overhead is estimated as 50% of the subtotal conversion costs less utilities, which include the projected costs for labor, maintenance, and analyses. Adminis'rative overhead is estimated as 10% of operating labor and supervision. Marketing the byproduct is also a consideration. Marketing is estimated as 10% of sales revenue.

Spent Catalyst Disposal--

For the dry FGT processes, the cost of disposing of the spent catalyst is added to the annual revenue requirement. For the Kurabo Knorca, UOP SFGT-N, and UOP SFGT-SN processes, disposal cost for the catalyst carriers (TiO_2, alumina, alumina, respectively) is used as the basis for the yearly disposal charge. For the Hitachi Zosen process, the catalyst base support has scrap metal value. Thus, this scrap value yields a credit toward the annual revenue requirement.

Byproduct Sales--

In the evaluation of annual and lifetime economics, credit from sale of byproducts is deducted from the yearly operating cost to give the net effect of the FGT process on the cost of power. Since byproduct credits are provided only for sulfuric acid in this study, the UOP SFGT-SN simultaneous SO_x-NO_x control process is the only system given a byproduct credit. A 1980 projected sales price for the byproduct sulfuric acid from the UOP process was estimated at $30/ton.

SYSTEMS ESTIMATED

Process descriptions, flowsheets, material balances, and major equipment lists and descriptions were prepared for each of the eight FGT processes included in this evaluation. Each process was divided into major operational areas to facilitate investment and revenue requirement comparisons for similar processing steps. Using the material balances, each piece of equipment was described, sized, costed, and listed by processing subsection, with total equipment costs in 1979 dollars for each item. The additional items of cost in each area which were not shown in the equipment lists, (i.e., installation labor and material costs for electrical, piping, ductwork, foundations, structural, instrumentation, insulation, and site preparation) were incorporated in the area investment estimates shown in the summary capital investment estimates in the following section.

CLASSIFICATION OF NO_x REMOVAL PROCESSES

The NO_x FGT processes can be divided into two types, dry or wet, based on whether or not the NO_x is absorbed into an aqueous solution. The primary methods of dry NO_x FGT which have been studied are: (1) catalytic decomposition, (2) SCR, (3) nonselective catalytic reduction, (4) selective noncatalytic reduction, (5) adsorption, and (6) electron beam radiation (6). About half of all the NO_x FGT processes currently undergoing development is the SCR type, which is the most widely tested and technically advanced type of FGT process. SCR is the only type of dry NO_x FGT process included in this study.

All current SCR-type processes use NH_3 to selectively reduce NO_x to molecular nitrogen and water in the presence of a catalyst. The reactions of NH_3 with NO_x are usually expressed by the process developers as one of the following:

$$4NH_{3(g)} + 6NO_{(g)} \rightarrow 5N_{2(g)} + 6H_2O_{(g)} \qquad (1)$$

$$8NH_{3(g)} + 6NO_{2(g)} \rightarrow 7N_{2(g)} + 12H_2O_{(g)} \qquad (2)$$

$$4NH_{3(g)} + 4NO_{(g)} + O_{2(g)} \rightarrow 4N_{2(g)} + 6H_2O_{(g)} \qquad (3)$$

$$4NH_{3(g)} + 2NO_{2(g)} + O_{2(g)} \rightarrow 3N_{2(g)} + 6H_2O_{(g)} \qquad (4)$$

Reactions (3) and (4) may more accurately represent the reduction reactions since several studies have demonstrated that the presence of at least some oxygen improves the NO_x-reduction efficiency (3, 4, 9, 10). As 90-95% of the total NO_x formed during the combustion of coal is present in the flue gas in the form of NO and NO_2 is only 5-10%, reaction (3) was used as the basis for material balance calculations in this study.

Process developers reported that most of the excess NH_3 reacts with oxygen as follows:

$$4NH_{3(g)} + 3O_{2(g)} \rightarrow 2N_{2(g)} + 6H_2O_{(g)} \qquad (5)$$

The NH_3 level in the treated flue gas leaving the FGT systems is low, reportedly less than 20 ppm.

There are some important factors affecting performance of these SCR processes which should be mentioned. The operating temperature range for most of these processes is about 300-400°C, though the more efficient NO_x removal usually occurs in the higher portion of this range. To maintain the reactor temperature at desirable operating levels during periods of reduced boiler load, most process vendors recommend bypassing a part of the flue gas around the economizer. In some pilot-plant and larger operations, auxiliary heaters have been used to maintain reactor temperatures during turndown. However, if this were necessary for large utility boilers, the economics of the SCR processes would be significantly increased.

The formation of ammonium bisulfate [$(NH_4)HSO_4$] and ammonium sulfate [$(NH_4)_2SO_4$] is another problem to consider. $(NH_4)HSO_4$ condensation usually occurs downstream of the reactor as the flue gases are cooled in the air heater. This presents potential corrosion problems and also reduced heat transfer problems. Also, some vendors have indicated problems with deposits of $(NH_4)_2SO_4$ in the reactor. Soot blowers for the reactor and air heater are the most commonly reported treatment of this situation. However, the information about this problem is still limited and the reported frequency of use of the soot blowers or other means varies greatly among vendors. Thus, an adequate evaluation of this problem is not possible here.

Another problem which may need to be addressed in future development programs is the monitoring of low concentrations of NH_3 and NO_x in the reactor outlet. Most vendors state that the NH_3 feed rate is partly controlled based on the NO_x and NH_3 concentrations in the flue gas leaving the reactor. There are fears by some, however, that monitoring of these low concentrations of NO_x and NH_3 may be difficult and that $(NH_4)HSO_4$ or $(NH_4)_2SO_4$ deposits in the sampling lines and instrument probes may compound the problem.

The dry SCR processes included in this report are offered by Hitachi Zosen; Kurabo Industries, Ltd.; and UOP, Inc. (UOP offers both the simultaneous SO_x-NO_x process and the NO_x-only process).

The wet FGT processes were, in most cases, designed to take advantage of technology already available from the previously developed FGD systems. Most of the wet processes were originally designed as simultaneous SO_x-NO_x systems. Unfortunately, the solubility of NO, composing 90-95% of the total NO_x present in coal-fired boiler flue gas, is low in aqueous solution and thus imposes a formidable obstacle to wet NO_x FGT processes. NO_2, which is the lesser component of the NO_x, is much more soluble than NO and does not impose as difficult a problem, although the solubility of NO_2 is also poor relative to SO_x. Therefore, the major task associated with any wet NO_x removal process is the absorption of the NO_x by the scrubbing solution where it can be concentrated and more economically converted into other forms.

There are two common methods of removing the NO_x from the flue gas: (1) direct absorption of the NO_x in the absorbing solution or (2) gas-phase oxidation to convert the relatively insoluble NO, either partially or entirely, to NO_2. Thus, the wet NO_x removal processes can be classified as absorption or oxidation processes, depending on whether or not the flue gas is treated with a gas-phase oxidant before absorption.

Additionally, each of these classifications can be divided based on the method by which the NO_x is converted after it has been absorbed by the scrubbing solution. Processes which reduce the absorbed NO_x either partially or completely to molecular nitrogen or complex nitrogen-sulfur compounds can be classified as reduction processes. Processes which do not reduce the absorbed NO_x are absorption processes. Thus, the wet NO_x processes can be categorized into one of the following groups: oxidation-absorption-reduction, oxidation-absorption, absorption-reduction, or absorption-oxidation. Figure 1 shows this classification system.

Because of the limited development status, technical feasibility, and also the doubtful economics of some of the process classifications, only two classifications of these wet NO_x FGT processes are evaluated in this report: oxidation-absorption-reduction and absorption-reduction processes. Included in the oxidation-absorption-reduction category is the Moretana Calcium process offered by Sumitomo Metal Industries and IHI process. The Asahi Chemical process is the example of absorption-reduction technology included in this study.

The oxidation-absorption-reduction processes are based on the use of a gas-phase oxidant to convert essentially all of the NO to NO_2. Before entering the absorber the flue gas is injected with an oxidant agent, either ozone (used by IHI) or ClO_2 (used in Moretana Calcium process).

This gas-phase oxidant rapidly and selectively oxidizes the NO to NO_2 which is then absorbed into an aqueous sulfite solution. The SO_x in the flue gas is simultaneously absorbed and forms either the sulfite or bisulfite ion in the scrubbing solution. The sulfite ions are partially oxidized to sulfate during reduction of the absorbed NO_2 and, depending on the process, the remaining sulfite may be oxidized by air to sulfate in the regeneration section. The sulfate is removed as gypsum. The absorbed NO_2 is at least partially reduced and removed as a combination of nitrate salts, molecular nitrogen, and complex nitrogen-sulfur compounds.

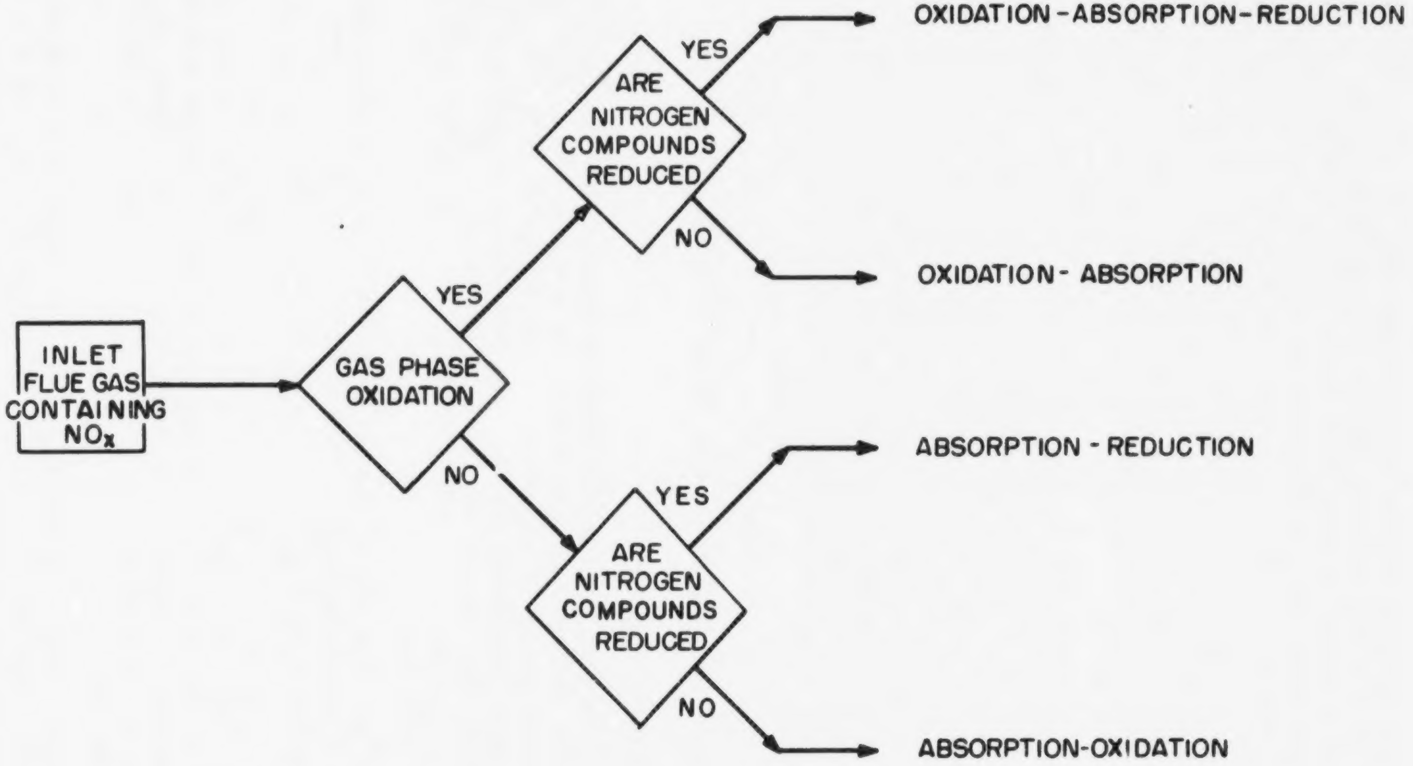
INLET FLUE GAS CONTAINING NO$_x$
GAS PHASE OXIDATION
YES
NO
ARE NITROGEN COMPOUNDS REDUCED
YES
NO
ARE NITROGEN COMPOUNDS REDUCED
YES
NO
OXIDATION - ABSORPTION - REDUCTION
OXIDATION - ABSORPTION
ABSORPTION - REDUCTION
ABSORPTION-OXIDATION

Figure 1. Classification system for wet NO$_x$ removal processes.

Because the Moretana Calcium process uses ClO_2 to oxidize NO to NO_2, approximately half of the absorbed NO_2 is reduced to molecular nitrogen and the other half is converted to nitrate salts. The IHI process, which uses ozone as the gas-phase oxidant, is based on limestone slurry scrubbing with a proprietary catalyst added to the scrubbing solution to enhance NO_2 absorption. This use of ozone and the addition of the catalyst apparently results in most of the NO_2 being reduced to complex nitrogen-sulfur compounds such as sulfamine salts. Only relatively minor amounts of molecular nitrogen and nitrate salts are formed in the process. These complex nitrogen-sulfur compounds are then thermally decomposed into molecular nitrogen and sulfate salts.

The absorption-reduction processes were specifically developed for the simultaneous removal of SO_x and NO_x from power plant flue gas without using an expensive gas-phase oxidant. These processes are based on certain ferrous-chelating compounds which improve the absorption of the relatively insoluble NO. The NO is absorbed by the scrubbing solution and forms a complex with the chelating compound. The SO_x is simultaneously absorbed by the scrubbing solution as the sulfite ion and reacts with the NO complex. The NO is reduced to molecular nitrogen and the ferrous-chelating compound is regenerated while the sulfite ion is being oxidized to sulfate. This sulfate is generally removed as gypsum.

ASAHI CHEMICAL SO_x-NO_x PROCESS - WET, ABSORPTION-REDUCTION

Process Description

This process developed by the Asahi Chemical Industry Company uses sodium sulfite (Na_2SO_3) solution to absorb both SO_x and NO_x (16). A ferrous-chelating compound is used to aid in the absorption of NO_x. The five main sections of this system are: (1) prescrubbing, (2) absorption of both SO_x and NO_x, (3) reduction of NO_x, (4) decomposition of dithionate, and (5) gypsum production. The Asahi process, as a whole, is an intricate system made up of many simple processing units.

The flue gas from the air heater is passed through a prescrubber which removes 99% of both particulates and chlorides. In addition, some of the scrubbing solution evaporates to humidify and to adiabatically cool the flue gas from 300°F to 127°F. The liquid prescrubber effluent is recirculated, makeup water is added, and small purge stream is removed and pumped to a fly ash thickener. The thickener overflow stream is recycled to the prescrubber and the bottoms containing the fly ash and the chlorides are removed, neutralized by limestone, and pumped to a waste disposal pond. Thus, the fly ash and chlorides in the flue gas are removed and treated before the flue gas reaches the absorber.

After leaving the prescrubber the flue gas enters a packed-bed absorber where it flows countercurrent to a 6.3 pH sodium-salt scrubbing solution containing ferrous ethylenediaminetetraacetic acid ($Fe^{+2}\cdot EDTA$). As the flue gas passes through the lower portion of the absorber, the SO_2 is rapidly absorbed into the solution and undergoes the following reactions:

$$SO_{2(g)} \rightarrow SO_{2(aq)} \tag{1}$$

$$Na_2SO_{3(aq)} + SO_{2(aq)} + H_2O \rightarrow 2NaHSO_{3(aq)} \tag{2}$$

$$2NaHSO_{3(aq)} + 1/2O_{2(aq)} \rightarrow Na_2S_2O_{6(aq)} + H_2O \tag{3}$$

The NO_x, on the other hand, is gradually absorbed over the entire length of the absorber and undergoes the following reactions to form ferrous-chelate complex:

$$NO_{(g)} \rightarrow NO_{(aq)} \tag{4}$$

$$NO_{(aq)} + Fe^{+2}\cdot EDTA_{(aq)} \rightarrow Fe^{+2}\cdot EDTA\cdot NO_{(aq)} \tag{5}$$

In addition to these primary reactions, other side reactions occur simultaneously. Both the $Fe^{+2}\cdot$EDTA and the Na_2SO_3 are readily oxidized in the scrubbing solution by oxygen absorbed from the flue gas to form ferric EDTA ($Fe^{+3}\cdot$EDTA) and Na_2SO_4:

$$O_{2(g)} \rightarrow O_{2(aq)} \tag{6}$$

$$Na_2SO_{3(aq)} + 1/2O_{2(aq)} \rightarrow Na_2SO_{4(aq)} \tag{7}$$

$$Fe^{+2}\cdot EDTA_{(aq)} \xrightarrow{O_2} Fe^{+3}\cdot EDTA_{(aq)} \tag{8}$$

The flue gas leaving the absorber is passed through a mist eliminator, reheated by mixing with the flue gas from the thermal decomposers, and sent to the stack.

The liquid effluent from the absorber is pumped to a reducing tank where makeup soda ash (Na_2CO_3) is added to replace sodium lost from the system and to convert sodium bisulfite ($NaHSO_3$) to Na_2SO_3. The reactions occurring in the reducing tank include:

$$Fe^{+2}\cdot EDTA\cdot NO_{(aq)} + Na_2SO_{3(aq)} \rightarrow Fe^{+2}\cdot EDTA_{(aq)} + Na_2SO_{4(aq)} + 1/2N_{2(g)} \tag{9}$$

$$Na_2CO_{3(aq)} + 2NaHSO_{3(aq)} \rightarrow 2Na_2SO_{3(aq)} + H_2O + CO_{2(g)} \tag{10}$$

Most of the NO absorbed is thus converted to molecular nitrogen and vented to the atmosphere. In addition, the $Fe^{+3}\cdot$EDTA which was produced in the absorption column (see equation 8) is reduced by the sulfite ion to $Fe^{+2}\cdot$EDTA.

Most of the effluent from the reducing tank is recycled to the absorber. The remainder (10-20% of the circulating solution) is pumped to a multiple-effect evaporator system in the regeneration section. The concentrated solution from the evaporators is then pumped to a cooling crystallizer where hydrated sodium dithionate ($Na_2S_2O_6$) and Na_2SO_4 crystals are produced under vacuum at 50°F. These crystals are separated from the mother liquor in a screw decanter and sent to a dryer operating at 250-300°F in which the hydrated crystals are converted to anhydrous sodium salts. For example, the reaction for the $Na_2S_2O_6$ crystals is:

$$Na_2S_2O_6 \cdot 2H_2O_{(s)} \xrightarrow{\Delta} Na_2S_2O_{6(s)} + 2H_2O_{(g)} \tag{11}$$

Most of the mother liquor from the decanter is recycled to the reducing tank and a smaller stream is passed through a sodium imidodisulfonate [$NH(SO_3Na)_2$] treatment section. This section is necessary since soluble $NH(SO_3Na)_2$ is formed in small quantities in the absorber and must be removed to prevent its buildup in the scrubbing solution. The $NH(SO_3Na)_2$ is converted to relatively insoluble potassium imidodisulfonate [$NH(SO_3K)_2$] by reaction with potassium sulfate (K_2SO_4):

$$NH(SO_3Na)_{2(aq)} + K_2SO_{4(aq)} \rightarrow NH(SO_3K)_{2(s)} + Na_2SO_{4(aq)} \tag{12}$$

The crystals of $NH(SO_3K)_2$ are separated in a screw decanter and sent to a thermal cracker. The mother liquor, mostly dissolved Na_2SO_4, is recycled to the reducing tank. The thermal cracker, which is heated indirectly by flue gas from an oil-fired furnace, operates at about 930°F and decomposes the $NH(SO_3K)_2$ into SO_2, molecular nitrogen, potassium sulfite (K_2SO_3), and K_2SO_4 by the following reaction:

$$2NH(SO_3K)_{2(s)} \underset{\Delta}{\rightarrow} 2SO_{2(g)} + K_2SO_{4(s)} + K_2SO_{3(s)} + H_2O_{(g)} + N_{2(g)} \tag{13}$$

The concentrated SO_2 gas stream is sent to the Na_2SO_4 converter while the solid K_2SO_3 and K_2SO_4 are recycled to the imidodisulfonate treatment section.

The anhydrous $Na_2S_2O_6$ and Na_2SO_4 mixture from the dryer is sent to a thermal cracker where it is heated indirectly to about 572°F to decompose the $Na_2S_2O_6$ by the following reaction:

$$Na_2S_2O_{6(s)} \underset{\Delta}{\rightarrow} Na_2SO_{4(s)} + SO_{2(g)} \tag{14}$$

A portion of this Na_2SO_4 is removed as a byproduct, but most is sent to a premixing tank where it is dissolved in makeup water and reacted with calcium sulfite ($CaSO_3$) to produce a slurry. This slurry is pumped to a Na_2SO_3 converter where it is treated with the SO_2 from both the dithionate and the imidodisulfonate thermal crackers to form gypsum and $NaHSO_3$ by the following reactions:

$$CaSO_{3(s)} + SO_{2(g)} + H_2O \rightarrow Ca(HSO_3)_{(aq)} \qquad (15)$$

$$Ca(HSO_3)_{2(aq)} + Na_2SO_{4(aq)} + 2H_2O \rightarrow 2NaHSO_{3(aq)} + CaSO_4 \cdot 2H_2O_{(s)} \qquad (16)$$

The converter product stream is pumped to a screw decanter where gyspum is removed as a solid byproduct. The liquor is sent to an SO_2 stripper, operating at 200°F, pH 4, and a pressure of 100 mmHg, where the bisulfite salts are partially converted to sulfites and SO_2:

$$Ca(HSO_3)_{2(aq)} \rightarrow CaSO_3 \cdot 1/2H_2O_{(s)}\downarrow + SO_{2(g)}\uparrow + 1/2H_2O \qquad (17)$$

$$2NaHSO_{3(aq)} \rightarrow Na_2SO_{3(aq)} + SO_{2(g)}\uparrow + H_2O \qquad (18)$$

The SO_2 produced in the stripper is recycled to the sodium sulfate converter and the effluent containing the sulfite is sent to the double decomposition reactor where the Na_2SO_3 is converted to $CaSO_3$ by the following reaction with makeup limestone slurry:

$$2NaHSO_{3(aq)} + CaCO_{3(s)} \rightarrow$$
$$Na_2SO_{3(aq)} + CaSO_3 \cdot 1/2H_2O_{(s)}\downarrow + 1/2H_2O + CO_{2(g)} \qquad (19)$$

The effluent from the double decomposition reactor is pumped to a thickener where the bottoms, a $CaSO_3$ slurry, are removed and pumped to a screw decanter. The solid from the decanter is then reslurried with makeup water and recycled to the sodium sulfate converter. The liquor from the decanter is recycled to the gypsum thickener. The thickener overflow, which is essentially a Na_2SO_3 solution, is sent to a regenerated solution tank, where makeup $Fe^{+2} \cdot$EDTA is added, and is then recycled to the reducing tank.

Analysis of Processing Subsections

The flow diagram and the material balance for the Asahi process are shown in Figure 2 and Table 25 respectively. To facilitate cost determinations and comparisons, the Asahi process was divided into 11 processing sections and the processing equipment was assigned to the appropriate section. The equipment list, giving the description and installed cost by section, is shown in Table 26.

Each of these processing sections is described in more detail below.

Material Handling--

This and the following area compose the limestone preparation section. The material-handling section includes all equipment needed to receive the limestone by rail or truck and to maintain a supply of limestone to the weigh feeders. One train of processing equipment is used.

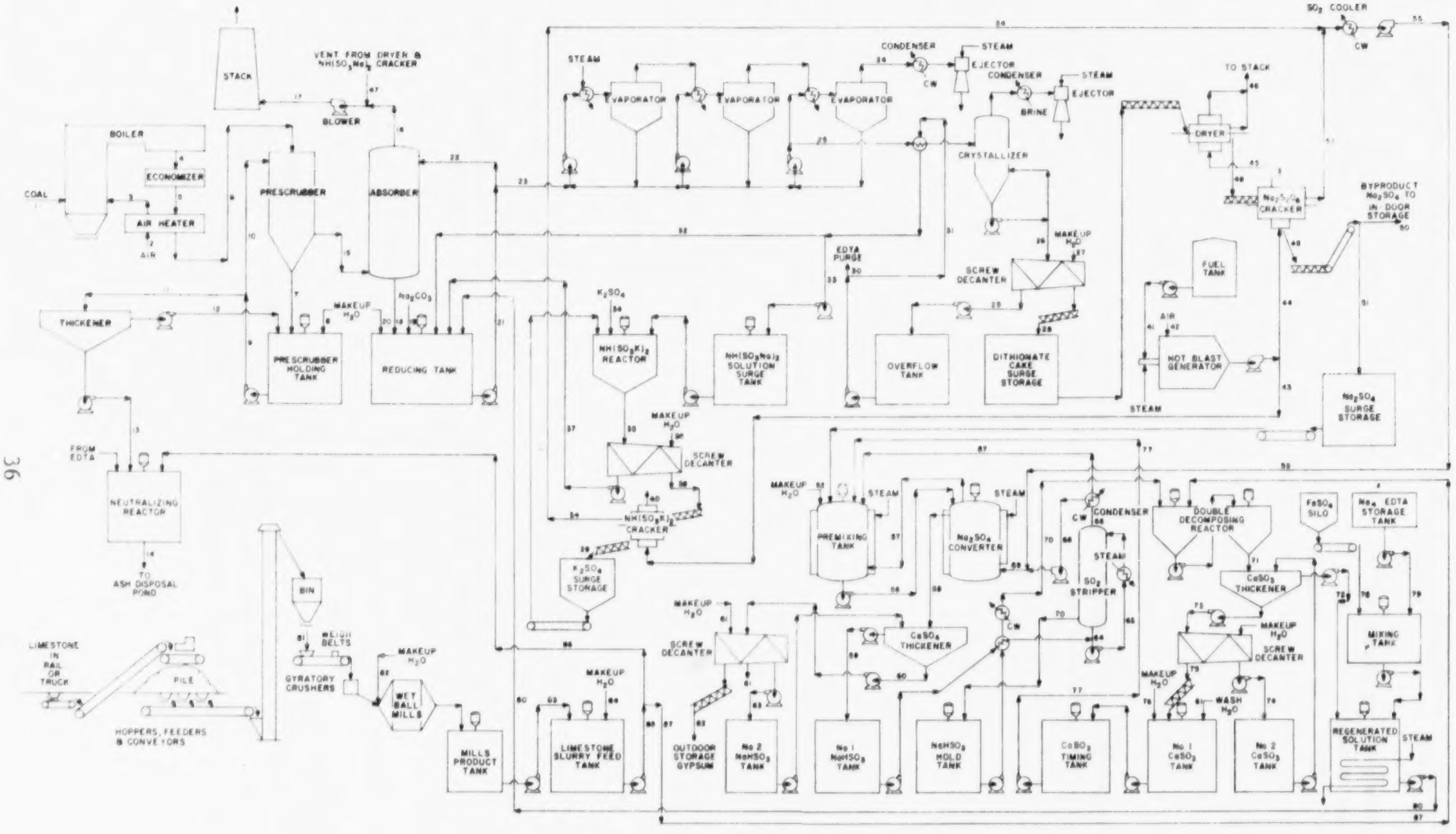
STACK
VENT FROM DRYER & NH(SO3Na)2 CRACKER
BLOWER
BOILER
COAL
ECONOMIZER
AIR HEATER
AIR
PRESCRUBBER
ABSORBER
THICKENER
PRESCRUBBER HOLDING TANK
MAKEUP H2O
Na2CO3
REDUCING TANK
FROM EDTA
NEUTRALIZING REACTOR
TO ASH DISPOSAL POND
LIMESTONE IN RAIL OR TRUCK
PILE
HOPPERS, FEEDERS & CONVEYORS
BIN
WEIGH BELTS
GYRATORY CRUSHERS
WET BALL MILLS
MILLS PRODUCT TANK
LIMESTONE SLURRY FEED TANK
STEAM
EVAPORATOR
EVAPORATOR
EVAPORATOR
CONDENSER
CW
EJECTOR
CONDENSER
BRINE
EJECTOR
CRYSTALLIZER
SCREW DECANTER
DITHIONATE CAKE SURGE STORAGE
OVERFLOW TANK
EDTA PURGE
K2SO4
NH(SO3K)2 REACTOR
NH(SO3Na)2 SOLUTION SURGE TANK
SCREW DECANTER
NH(SO3K)2 CRACKER
K2SO4 SURGE STORAGE
DRYER
TO STACK
Na2S2O6 CRACKER
SO2 COOLER
BYPRODUCT Na2SO4 TO IN DOOR STORAGE
FUEL TANK
HOT BLAST GENERATOR
Na2SO4 SURGE STORAGE
PREMIXING TANK
Na2SO4 CONVERTER
CONDENSER
SO2 STRIPPER
CaSO4 THICKENER
SCREW DECANTER
OUTDOOR STORAGE GYPSUM
No 2 NaHSO3 TANK
No 1 NaHSO3 TANK
NaHSO3 HOLD TANK
CaSO3 TIMING TANK
DOUBLE DECOMPOSING REACTOR
CaSO3 THICKENER
SCREW DECANTER
WASH H2O
No 1 CaSO3 TANK
No 2 CaSO3 TANK
FeSO4 SILO
Na4 EDTA STORAGE TANK
MIXING TANK
REGENERATED SOLUTION TANK

Figure 2. Asahi SO_x-NO_x process. Flowsheet.

TABLE 25. ASAHI SO_x-NO_x PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	428,600	4,546,200	4,101,800	4,516,100	4,516,100
2						
3	Sft^3/min (60°F)		1,005,000	906,700	958,000	958,000
4	Temperature, °F		80	535	890	705
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to prescrubber	Effluent from prescrubber	Makeup H_2O to prescrubber	Effluent from prescrubber holding tank	Recycle to top of prescrubber
1	Total stream, lb/hr	4,960,400	3,743,200	332,700	6,596,400	3,895,400
2						
3	Sft^3/min (60°F)	1,056,000				
4	Temperature, °F	300			127	127
5	Pressure, psig					
6	Gpm				12,546	7,415
7	Specific gravity		1.06	1.0	1.05	1.05
8	pH					
9	Undissolved solids, %		3.5		2	2
10						

	Stream No.	11	12	13	14	15
	Description	Thickener feed	Thickener overflow to prescrubber holding tank	Thickener bottoms to neutralization reactor	Ash slurry purge to disposal pond	Flue gas from prescrubber to absorber
1	Total stream, lb/hr	2,701,000	2,520,600	180,400	206,500	5,113,000
2						
3	Sft^3/min (60°F)					1,127,875
4	Temperature, °F	127				127
5	Pressure, psig					
6	Gpm	5,131				
7	Specific gravity	1.05			1.26	
8	pH					
9	Undissolved solids, %	2			30.2	
10						

	Stream No.	16	17	18	19	20
	Description	Flue gas from absorber to reheat	Flue gas to the stack	Effluent from absorber	Makeup soda ash	Makeup H_2O
1	Total stream, lb/hr	5,084,300	5,322,800	48,868,200	9,750	33,500
2						
3	Sft^3/min (60°F)	1,124,079				
4	Temperature, °F	127				
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

(continued)

TABLE 25. (continued)

	Stream No.	21	22	23	24	25
	Description	Effluent from reducing tank	Absorber recycle	Triple effect evaporator feed	Condenser feed	Crystallizer feed
1	Total stream, lb/hr	48,840,000	47,630,000	1,210,000	25,000	1,073,900
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	26	27	28	29	30
	Description	Forward feed to $Na_2S_2O_6$ separator	Wash H_2O to $Na_2S_2O_6$ separator	$Na_2S_2O_6$ cake surge storage tank feed	Centrate feed to overflow tank	Bleed to EDTA recovery
1	Total stream, lb/hr	1,048,900	6,400	126,600	922,300	9,900
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity	1.69				
8	pH					
9	Undissolved solids, %	11.6				
10						

	Stream No.	31	32	33	34	35
	Description	Centrate recycle	Centrate recycle to reducing tank	Centrate feed to imidodisulfonate section	Makeup K_2SO_4	$NH(SO_3K)_2$ separator feed
1	Total stream, lb/hr	914,400	854,800	57,500	300	62,100
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	36	37	38	39	40
	Description	Makeup wash H_2O	Centrate recycle to reducing tank	$NH(SO_3K)_2$ to cracker	Recycle K_2SO_3-K_2SO_4 from surge storage	Off-gas from $NH(SO_3K)_2$ cracker
1	Total stream, lb/hr	340	55,600	6,820	4,250	12,790
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

(continued)

TABLE 25. (continued)

	Stream No.	41	42	43	44	45
	Description	Fuel oil to furnace	Combustion air to furnace (15% excess O_2)	Flue gas to $NH(SO_3K)_2$ cracker	Flue gas to $Na_2S_2O_6$ cracker	Flue gas to dryer
1	Total stream, lb/hr	11,450	181,000	12,790	179,500	179,500
2						
3	Sft^3/min (60°F)					
4	Temperature, °F		80			
5	Pressure, psig					
6	Gpm	25				
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	46	47	48	49	50
	Description	Flue gas from dryer	Flue gas from dryer and $NH(SO_3K)_2$ cracker	$Na_2S_2O_6$ feed to cracker	Na_2SO_4 from cracker	Na_2SO_4 to indoor storage
1	Total stream, lb/hr	225,700	238,500	80,600	64,000	10,400
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	51	52	53	54	55
	Description	Na_2SO_4 to surge storage	Makeup H_2O to premixing tank	SO_2-rich off-gas from $Na_2S_2O_6$ cracker	SO_2-rich off-gas from $NH(SO_3K)_2$ cracker	Combined cracker off-gas
1	Total stream, lb/hr	53,600	31,600	16,600	2,575	19,200
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	56	57	58	59	60
	Description	Effluent from premixing tank	Converter off-gas recycled to premixing tank	Na_2SO_4 converter effluent to $CaSO_4$ thickener	Thickener overflow to No.1 $NaHSO_3$ tank	Thickener underflow to gypsum centrifuge
1	Total stream, lb/hr	278,100	2,200	299,800	253,500	159,700
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %	16.5				27.5
10						

(continued)

TABLE 25. (continued)

	Stream No.	61	62	63	64	65
	Description	Wash H_2O to gypsum centrifuge	Byproduct gypsum	Gypsum centrate to No. 2 $NaHSO_3$ tank	SO_2 stripper bottoms to recycle	Recycle stream to the top of SO_2 stripper
1	Total stream, lb/hr	2,310	46,200	113,500	1,675,900	1,789,400
2						
3	Sft³/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %		95			
10						

	Stream No.	66	67	68	69	70
	Description	SO_2 stripper off-gas	H_2O condensed from SO_2 stripper off-gas	SO_2 off-gas from condenser	Combined SO_2 off-gas streams to Na_2SO_4 converter	Effluent from SO_2 stripper
1	Total stream, lb/hr	50,700	42,200	8,460	27,610	202,800
2						
3	Sft³/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	71	72	73	74	75
	Description	Effluent from double decomposition reactor	Thickener overflow to regenerated solution tank	Thickener underflow to $CaSO_3$ centrifuge	Centrate to No. 2 $CaSO_3$ tank	Solids from centrifuge to No. 1 $CaSO_3$ tank
1	Total stream, lb/hr	280,700	227,300	64,800	11,400	53,400
2						
3	Sft³/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %			70		85
10						

	Stream No.	76	77	78	79	80
	Description	Makeup H_2O to No. 1 $CaSO_3$ tank	Makeup $CaSO_3$ slurry to premix tank	Makeup $FeSO_4$	Makeup EDTA	Regenerated solution to reducing tank
1	Total stream, lb/hr	95,600	151,300	240	250	227,600
2						
3	Sft³/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %		30			
10						

(continued)

TABLE 25. (continued)

	Stream No.	81	82	83	84	85
	Description	Makeup limestone from pile	Makeup H_2O to ball mill	Makeup limestone slurry to limestone slurry feed tank	Makeup H_2O to limestone slurry feed tank	Effluent from limestone slurry feed tank
1	Total stream, lb/hr	32,100	21,400	53,500	53,500	107,000
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity			1.61		
8	pH					
9	Undissolved solids, %			60		15
10						

	Stream No.	86	87			
	Description	Makeup limestone slurry to neutralization reactor	Makeup limestone slurry to double decomposition reactor			
1	Total stream, lb/hr	19,800	87,200			
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %	15	15			
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

TABLE 26. ASAHI SO_x-NO_x PROCESS

BASE CASE EQUIPMENT LIST
DESCRIPTION AND COST

Area 1--Material Handling

	Item	No.	Description	Total equipment cost, 1979 $
1.	Car shaker and hoist	1	20 hp shaker, 7.5 hp hoist	31,000
2.	Car puller	1	25 hp puller with 5 hp return	52,000
3.	Hopper, unloading	1	16 ft dia x 10 ft straight side height, carbon steel	12,800
4.	Feeder, unloading vibrating	1	3.5 hp	14,400
5.	Conveyor, unloading belt	1	20 ft horizontal, 5 hp	17,700
6.	Conveyor, unloading incline belt	1	310 ft, 50 hp	89,200
7.	Unloading pit dust collector	1	Polypropylene bag type, 2,200 ft^3/min, 7.5 hp	19,200
8.	Pump, unloading pit sump	1	60 gpm, 70 ft head, 5 hp	4,200
9.	Conveyor, storage belt	1	100 ft, 2 hp	45,200
10.	Tripper, storage conveyor	1	30 ft^3/min, 1 hp	16,400
11.	Mobile equipment	1	Scraper tractor	137,800
12.	Hopper, reclaim	1	7 ft wide, 4.25 ft high x 2 ft wide bottom, carbon steel	1,500
13.	Feeder, reclaim vibrating	1	3.5 hp	14,400
14.	Conveyor, reclaim belt	1	100 ft, 2 hp	32,200

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
15. Conveyor, reclaim incline belt	1	193 ft, 4 hp	53,800
16. Reclaim pit dust collector	1	Polypropylene bag type	19,200
17. Pump, reclaim pit sump	1	60 gpm, 70 ft head, 5 hp	4,200
18. Elevator, reclaim bucket	1	90 ft high, 75 hp	86,400
19. Bin, feed	2	13 ft dia x 21 ft straight side height, covered, carbon steel	33,200
Subtotal			684,800

Area 2--Feed Preparation

Item	No.	Description	Total equipment cost, 1979 $
1. Feeder, bin weigh	2	14 ft pulley centers, 2 hp	39,600
2. Gyratory crushers	2	75 hp	119,500
3. Ball mill dust collectors	2	Polypropylene bag type, 2,200 ft^3/min, 7.5 hp	38,400
4. Ball mill	2	9.3 tons/hr, 125 hp	320,100
5. Tank, ball mill product	2	10 ft dia x 10 ft height, 5,500 gal, open top, flakeglass-lined, carbon steel	26,800
6. Agitator, mills product tank	2	10 hp, resin-lined, carbon steel	17,500

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
7. Pump, ball mill product tank	3	Centrifugal belt drive, 55 gpm, 50 ft head, 5 hp resin-lined, carbon steel	7,800
8. Tank, slurry surge	1	40 ft dia x 45 ft height, 42,300 gal, open top, resin-lined, carbon steel	181,700
9. Pump, limestone slurry feed	2	Centrifugal belt drive, 220 gpm, 100 ft head, 12 hp, resin-lined, carbon steel	11,100
10. Agitator, limestone slurry tank	1	75 hp, resin-lined, carbon steel	68,400
Subtotal			830,900

Area 3--Gas Handling

Item	No.	Description	Total equipment cost, 1979 $
1. Blowers, stack gas	4	388,200 ft^3/min, ΔP, 39.4 in. H_2O, 1,500 hp, carbon steel	2,269,400
Subtotal			2,269,400

Area 4--Particulate Control

Item	No.	Description	Total equipment cost, 1979 $
1. Prescrubbers	4	Venturi, 1.97 ft wide x 32.8 ft long x 32.8 ft high, resin-lined, carbon steel	703,200
2. Tanks, prescrubber recycle	4	18 ft dia, 21.5 ft high, 40,900 gal, open top, resin-lined, carbon steel	147,200

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
3. Agitator, prescrubber holding tank	4	15 hp, resin-lined, carbon steel	86,400
4. Pump, prescrubber recycle	6	Centrifugal belt drive, 3,200 gpm, 144 ft head, 200 hp, resin-lined, carbon steel	109,500
5. Thickener, ash	1	82 ft dia x 13.1 ft high, 527,901 gal, resin-lined, carbon steel	574,300
6. Pump, thickener overflow recycle	2	Centrifugal belt drive, 5,100 gpm, 50 ft head, 125 hp, resin-lined, carbon steel	51,200
7. Pump, thickener underflow	2	Centrifugal belt drive, 300 gpm, 50 ft head, 10 hp, resin-lined, carbon steel	10,100
8. Reactor, ash neutralization	1	11.5 ft dia x 14.8 ft high, 11,500 gal, open top, resin-lined, carbon steel	16,100
9. Agitator, neutralization reactor	1	6 hp, resin-lined, carbon steel	13,000
10. Pump, ash to pond	2	Centrifugal belt drive, 350 gpm, 300 ft head, 60 hp, resin-lined, carbon steel	32,500
Subtotal			1,743,500

Area 5--SO_x-NO_x Absorption

Item	No.	Description	Total equipment cost, 1979 $
1. Absorbers	4	Tellerette packed column, 36 ft long x 36 ft wide x 33 ft high, 316 SS	8,816,900

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
2. Tanks, reducing	4	36 ft dia x 36 ft high, 274,100 gal, closed top, resin-lined, carbon steel	623,800
3. Agitator, reducing tank	4	50 hp, resin-lined, carbon steel	203,100
4. Pump, absorber recycle	6	Centrifugal belt drive, 15,700 gpm, 135 ft head, 1,000 hp, resin-lined, carbon steel	353,500
Subtotal			9,997,300

Area 6--Crystallizing

Item	No.	Description	Total equipment cost, 1979 $
1. Evaporator	2	Triple effect, 24 ft dia x 19.7 ft high, 66,000 gal, 316 SS	1,110,000
2. Pump, recycle first effect evaporator	2	Centrifugal belt drive, 4,600 gpm, 75 ft head, 200 hp, rubber-lined, carbon steel	39,000
3. Pump, recycle second effect evaporator	2	Centrifugal belt drive, 4,300 gpm, 75 ft head, 200 hp, rubber-lined, carbon steel	39,000
4. Pump, recycle third effect evaporator	2	Centrifugal belt drive, 4,000 gpm, 100 ft head, 200 hp, rubber-lined, carbon steel	39,000
5. Condenser, third effect evaporator	2	5,300 ft^2, 316 SS	472,100

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
6. Steam ejector, third effect evaporator	2	100 psig, steam consumption 1,100 lb/hr, 1-1/2 in. syphon, 90°F suction temperature, 20 ft suction lift, resin-lined, carbon steel	5,200
7. Heat exchanger, feedstream to crystallizer	2	3,000 ft^2, 316 SS	132,000
8. Crystallizer, cooling	2	16.4 ft dia x 23 ft high, 37,000 gal, resin-lined, carbon steel	73,100
9. Condenser, at crystallizer	1	25,000 ft^2, air cooled, 316 SS	1,947,000
10. Refrigeration system	1	Cooling load, 4.87 x 10^6 kcal/hr	270,000
11. Steam ejector - crystallizer	2	100 psig, steam consumption 2,200 lb/hr, 1-1/2 in. syphon, 41°F suction temperature, 20 ft suction lift, resin-lined, carbon steel	10,400
12. Pump, crystallizer recycle	4	Centrifugal belt drive, 3,700 gpm, 75 ft head, 200 hp, resin-lined, carbon steel	78,000
13. Separator, dithionate	13	Screw decanter, 316 SS	2,225,500
14. Pump, overflow to surge tank	2	Centrifugal belt drive, 1,100 gpm, 75 ft head, 40 hp, resin-lined, carbon steel	18,100
15. Tank, centrate overflow surge	1	35 ft dia x 37 ft high, 266,300 gal, open top, resin-lined, carbon steel	129,100
16. Conveyor, dithionate cake to storage	1	Screw, 16 in. dia x 30 ft long, 40 hp, 65 tons/hr	10,800

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
17. Tank, dithionate cake surge	1	30 ft dia x 30 ft high, 158,600 gal, open top, resin-lined, carbon steel	91,400
18. Pump, overflow to reduction reactor	2	Centrifugal belt drive, 1,100 gpm, 150 ft head, 100 hp, resin-lined, carbon steel	24,800
19. Building	1	60 ft x 120 ft, 1 level	230,400
Subtotal			6,944,900

Area 7--Nitrogen Treatment

Item	No.	Description	Total equipment cost, 1979 $
1. Pump, $NH(SO_3Na)_2$ solution	2	Centrifugal belt drive, 70 gpm, 50 ft head, 2 hp, resin-lined, carbon steel	4,800
2. Tank, $NH(SO_3Na)_2$ solution surge	1	18 ft dia x 18 ft high, 34,300 gal, open top, resin-lined, carbon steel	31,800
3. Agitator, $NH(SO_3Na)_2$ solution surge tank	1	5 hp, resin-lined, carbon steel	23,000
4. Pump, $NH(SO_3Na)_2$ reactor feed	2	Centrifugal belt drive, 70 gpm, 50 ft head, 2 hp, resin-lined, carbon steel	4,800
5. Reactor, $NH(SO_3Na)_2$	1	6.6 ft dia x 14.8 ft high, 3,800 gal, 316 SS	13,200
6. Agitator, $NH(SO_3Na)_2$ reactor	1	15 hp, resin-lined, carbon steel	9,900
7. Separator, $NH(SO_3Na)_2$	1	Screw decanter, 316 SS	171,200

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
8. Pump, recycle centrate to reducing tank	2	Centrifugal belt drive, 70 gpm, 75 ft head, 3 hp, resin-lined, carbon steel	4,800
9. Conveyor, $NH(SO_3Na)_2$ cake to cracker	1	Screw, 6 in. dia x 30 ft long, 7.5 hp, 3.5 tons/hr	4,400
10. Cracker, $NH(SO_3Na)_2$	1	Rotary kiln, indirect heating, 4.6 ft dia x 36 ft long, 33 kW, 310 SS	218,300
11. Conveyor, sulfate to storage	1	Screw, 6 in. dia x 30 ft long, 5 hp, 2.5 tons/hr	3,200
12. Tank, K_2SO_4 surge storage	1	10 ft dia x 10 ft high, 59,000 gal, open top, resin-lined, carbon steel	9,900
13. Weigh feeder, K_2SO_4 storage	1	4,300 lb/hr	6,400
14. Conveyor, sulfate to $NH(SO_3Na)_2$ reactor	1	Bucket, 6 in. x 4 in. x 4 in., 0.5 hp, 50 ft lift, 2.2 tons/hr	8,700
15. Tank, H_2SO_4 storage	1	16 ft dia x 20 ft high, 30,100 gal, closed top, carbon steel	19,800
16. Pump, H_2SO_4	2	Centrifugal belt drive, 40 gpm, 50 ft head, 2 hp, carbon steel	3,600
17. Reactor, EDTA	1	7.5 ft dia x 10 ft high, 1,075 gal, open top, resin-lined, carbon steel	7,100
18. Agitator-EDTA reactor	1	3 hp, resin-lined, carbon steel	6,700
19. Filter, EDTA	1	40 ft^2 filtering area	50,000
20. Pump, filtrate to neutralization reactor	2	Centrifugal belt drive, 100 gpm, 50 ft head, 2.5 hp, resin-lined, carbon steel	6,500
Subtotal			608,100

(continued)

TABLE 26. (continued)

Area 8--Cracking

	Item	No.	Description	Total equipment cost, 1979 $
1.	Tank, fuel oil storage	1	35 ft dia x 40 ft high, 287,900 gal, closed top, carbon steel	91,000
2.	Pump, fuel oil	2	Centrifugal belt drive, 25 gpm, 100 ft head, 1 hp, cast iron	2,800
3.	Hot blast generator (furnace)	1	35 x 10^6 kcal/hr, oil-fired	955,400
4.	Blower, flue gas to cracker	1	100,750 ft^3/min, 400 hp, ΔP 15 in. H_2O	208,500
5.	Conveyor, dithionate cake to dryer	1	Bucket, 12 in. x 7 in. x 7-1/4 in., 10 hp, 50 ft lift, 65 tons/hr	13,800
6.	Weigh feeder; dithionate surge to dryer	1	63.3 tons/hr	15,100
7.	Dryer	1	Rotary kiln, direct heating, 9.8 ft dia x 115 ft long, 66 kW, 304 SS	2,389,500
8.	Conveyor, dithionate cake to cracker	1	Screw, 12 in. dia x 30 ft long, 60 hp, 40.3 tons/hr	5,800
9.	Cracker, dithionate	7	Rotary kiln, indirect heating, 4.6 ft dia x 36 ft long, 33 kW, 310 SS	3,651,500
10.	Conveyor, Na_2SO_4 to storage	1	Screw, 14 in. dia x 30 ft long, 60 hp, 32 tons/hr	5,800
11.	Conveyor, Na_2SO_4 to storage	1	Belt, 14 in. wide x 100 ft long, 5 hp, 20 ft lift, 32 tons/hr	24,300

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
12. Tank, Na_2SO_4 surge storage	1	10 ft dia x 10 ft high, 5,900 gal, open top, resin-lined, carbon steel	11,800
13. Conveyor, byproduct Na_2SO_4 to storage	1	Belt, 14 in. wide x 300 ft long, 1 hp, 5.2 tons/hr	68,800
14. Tank, Na_2SO_4 byproduct storage	1	24 ft dia x 24 ft high, 81,200 gal, open top, resin-lined, carbon steel	68,700
Subtotal			7,512,800

Area 9--Gypsum Production

Item	No.	Description	Total equipment cost, 1979 $
1. Heat exchanger, SO_2 gas	1	500 ft^2, 316 SS	42,200
2. Blower, SO_2 No. 1	2	2,075 ft^3/min, 10 hp, ΔP 15 in. H_2O	3,300
3. Converter, Na_2SO_4	1	16.4 ft dia x 20 ft high, 31,700 gal, closed top, resin-lined, carbon steel	75,200
4. Agitator, Na_2SO_4 converter	1	30 hp, resin-lined, carbon steel	34,100
5. Conveyor, Na_2SO_4 to premix tank	1	Bucket, 8 in. x 5 in. x 5-1/2 in., 5 hp, 40 ft lift, 27 tons/hr	13,700
6. Weigh feeder, $NaSO_4$ to premix tank	1	26.8 tons/hr	11,000
7. Tank, premixing	1	27.5 ft dia x 27.5 ft wide, 122,200 gal, open top, resin-lined, carbon steel	163,000

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
8. Agitator, premix tank	1	60 hp, resin-lined, carbon steel	58,000
9. Pump, Na_2SO_4 converter feed	2	Centrifugal belt drive, 360 gpm, 50 ft head, 10 hp, resin-lined, carbon steel	10,400
10. Thickener, gypsum	1	75.5 ft dia x 16.4 ft high, 49,100 gal, resin-lined, carbon steel	380,200
11. Pump, gypsum separator feed	2	Centrifugal belt drive, 185 gpm, 50 ft head, 7.5 hp, resin-lined, carbon steel	7,200
12. Separator, gypsum	6	Screw decanter, 316 SS	578,400
13. Conveyor, gypsum to storage	1	Belt, 14 in. wide x 300 ft long, 7.5 hp, 50 ft lift, 23.2 tons/hr	76,600
14. Pump, $NaHSO_3$ tank No. 2 feed	2	Centrifugal belt drive, 140 gpm, 50 ft head, 5 hp, resin-lined, carbon steel	6,900
15. Tank, $NaHSO_3$ surge No. 2	1	15 ft dia x 15 ft high, 19,800 gal, open top, resin-lined, carbon steel	22,500
16. Pump, $NaHSO_3$ recycle	2	Centrifugal belt drive, 150 gpm, 50 ft head, 5 hp, resin-lined, carbon steel	6,900
17. Pump, $NaHSO_3$ tank No. 1 feed	2	Centrifugal belt drive, 325 gpm, 50 ft head, 7.5 hp, resin-lined, carbon steel	10,300
18. Tank, $NaHSO_3$ surge No. 1	1	18 ft dia x 21 ft high, 40,000 gal, open top, resin-lined, carbon steel	37,400
19. Pump, $NaHSO_3$ feed to SO_2 stripper	2	Centrifugal belt drive, 325 gpm, 100 ft head, 15 hp, resin-lined, carbon steel	11,500
Subtotal			1,548,800

(continued)

TABLE 26. (continued)

Area 10--Absorbent Regeneration

	Item	No.	Description	Total equipment cost, 1979 $
1.	Heat exchanger, SO_2 stripper feed	1	1,700 ft^2, 316 SS	70,600
2.	Pump, SO_2 stripper recycle	2	Centrifugal belt drive, 965 gpm, 100 ft head, 316 SS	16,500
3.	Heat exchanger, SO_2 stripper heater	1	1,400 ft^2, 316 SS	63,800
4.	SO_2 stripper	1	16.4 ft dia x 26.2 ft high, 39,600 gal, 316 SS	93,600
5.	Heat exchanger, SO_2 stripper condenser	1	700 ft^2, 316 SS	46,800
6.	Tank, $NaHSO_3$ holding	1	32.5 ft dia x 35 ft high, 217,200 gal, open top, resin-lined, carbon steel	111,200
7.	Agitator, $NaHSO_3$ holding tank	1	50 hp, resin-lined, carbon steel	49,300
8.	Pump, $NaHSO_3$ feed to double decomposition	2	Centrifugal belt drive, 260 gpm, 100 ft head, 20 hp, resin-lined, carbon steel	11,200
9.	Heat exchanger	1	400 ft^2, 316 SS	40,000
10.	Reactor, decomposition	2	16.4 ft dia x 19 ft high, 29,000 gal, open top, 304 SS	166,900
11.	Agitator, double decomposition reactor	2	25 hp, resin-lined, carbon steel	60,000
12.	Pump, reactor No. 2 feed	1	Centrifugal belt drive, 226 gpm, 50 ft head, 10 hp, resin-lined, carbon steel	8,600

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
13. Thickener, sulfite	1	75.4 ft dia x 16.4 ft high, 527,900 gal, resin-lined, carbon steel	375,300
14. Pump, regenerated solution tank feed	2	Centrifugal belt drive, 355 gpm, 50 ft head, 10 hp, resin-lined, carbon steel	10,500
15. Pump, $CaSO_3$ separator feed	2	Centrifugal belt drive, 75 gpm, 50 ft head, 3 hp, resin-lined, carbon steel	6,800
16. Separator, sulfite	6	Screw decanter, 304 SS	578,400
17. Conveyor, $CaSO_3$ cake to tank	1	Bucket, 8 in. x 5 in. x 5-1/2 in., 5 hp, 40 ft lift, 27 tons/hr	9,700
18. Tank, $CaSO_3$ No. 1	1	25 ft dia x 20 ft high, 102,800 gal, open top, resin-lined, carbon steel	68,100
19. Agitator, $CaSO_3$ tank No. 1	1	25 hp, resin-lined, carbon steel	30,000
20. Pump, $CaSO_3$ feed No. 1	2	Centrifugal belt drive, 265 gpm, 50 ft head, 7.5 hp, resin-lined, carbon steel	10,300
21. Tank, $CaSO_3$ timing	1	26 ft dia x 30 ft high, 119,100 gal, open top, resin-lined, carbon steel	75,500
22. Agitator, $CaSO_3$ timing, tank	1	25 hp, resin-lined, carbon steel	30,000
23. Pump, $CaSO_3$ feed No. 2	2	Centrifugal belt drive, 265 gpm, 150 ft head, 25 hp, resin-lined, carbon steel	11,800
24. Pump, No. 2 $CaSO_3$ tank feed	2	Centrifugal belt drive, 20 gpm, 50 ft head, 1 hp, resin-lined, carbon steel	4,800

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
25. Tank, $CaSO_3$ No. 2	1	6 ft dia x 10 ft high, 2,100 gal, open top, resin-lined, carbon steel	5,700
26. Pump, $CaSO_3$ recycle	2	Centrifugal belt drive, 20 gpm, 50 ft head, 1 hp, resin-lined, carbon steel	4,800
27. Tank, regenerated solution	1	45 ft dia x 47 ft high, 559,100 gal, open top, resin-lined, carbon steel	214,900
28. Agitator, solution tank	2	50 hp, resin-lined, carbon steel	98,700
29. Pump, regenerated solution	2	Centrifugal belt drive, 355 gpm, 100 ft head, 20 hp, resin-lined, carbon steel	11,200
30. Blower, SO_2	2	1,080 ft^3/min, 5 hp, ΔP 15 in. H_2O	3,300
Subtotal			2,288,300

Area 11--Raw Material Storage

Item	No.	Description	Total equipment cost, 1979 $
1. Elevator to storage bin	1	Bucket, 8 in. x 5 in. x 5-1/2 in., 10 hp, 80 ft lift, 30 tons/hr	15,400
2. Tank, Na_2CO_3 storage	1	Cylindrical tank with cone bottom, 37.5 ft dia x 40 ft high, 50,254 ft^3, closed top, carbon steel	184,900
3. Weigh feeder, Na_2CO_3 to reducing tank	1	9,750 lb/hr	6,400

(continued)

TABLE 26. (continued)

Item	No.	Description	Total equipment cost, 1979 $
4. Conveyor, Na_2CO_3 to reducing tank	1	Belt, 14 in. wide x 300 ft long, 2.5 hp, 40 ft lift, 5 tons/hr	77,000
5. Elevator, $FeSO_4$ to storage	1	Bucket, 6 in. x 4 in. x 4-1/2 in., 2.5 hp, 30 ft lift, 14 tons/hr	6,900
6. Hopper, $FeSO_4$ storage	1	Square hopper with pyramidal bottom, 10 ft long x 10 ft wide x 10 ft high, 1,433 ft^3, carbon steel	19,200
7. Weigh feeder, $FeSO_4$ to mix tank	1	238 lb/hr	5,600
8. Tank, Na_4EDTA storage	1	Cylindrical tank, 19 ft dia x 20 ft high, 42,400 gal, closed top, resin-lined, carbon steel	44,100
9. Pump, Na_4EDTA to mixing tank	2	Centrifugal belt drive, 60 gpm, 50 ft head, 2 hp, resin-lined, carbon steel	4,800
10. Mixing tank, Na_4EDTA-$FeSO_4$	1	5 ft dia x 5 ft high, 750 gal, open top, resin-lined, carbon steel	2,500
11. Agitator, Na_4EDTA-$FeSO_4$ mixing tank	1	3 hp, resin-lined, carbon steel	6,900
12. Pump, Fe·EDTA to reg. solution tank	2	Centrifugal belt drive, 60 gpm, 50 ft head, 2 hp, resin-lined, carbon steel	4,800
Subtotal			378,500
Total, Areas 1-11			34,807,300

Feed Preparation--

The feed preparation section consists of a single train of equipment required to weigh, crush, and slurry the makeup limestone. The fresh limestone slurry, containing 30 weight percent solids, is stored in the limestone slurry feed tank which has a surge volume sufficient for 24 hours normal operation.

Gas Handling--

Four ID fans, one for each train of prescrubbers and absorbers, are provided between the reheating section and the stack to compensate for the pressure drop in the Asahi system. Each fan is sized to handle one-fourth of the total flue gas volume and a pressure drop of 59 inches water in the system.

Particulate Control--

Four trains of modified, variable-throat venturi prescrubbers and four trains of prescrubber holding tanks are used. One train of equipment required to concentrate, neutralize, and pump the fly ash slurry to the pond is provided.

SO_x-NO_x Absorption--

Four trains of packed-bed absorbers and four reducing tanks are provided. Six recycle pumps (four operating and two spares) are provided to recirculate the scrubbing solution from the reducing tanks to the absorbers.

Crystallization Section--

Two triple-effect evaporator systems to concentrate the absorber purge solution and two cooling crystallizers are provided.

Na_2SO_4 and $Na_2S_2O_6$ crystals are separated in 13 screw decanters operating in parallel. The solids storage tank has an 8-hour capacity and the liquid storage tank has a 4-hour capacity.

Nitrogen Treatment Section--

A single train of processing equipment is used to separate and process $NH(SO_3Na)_2$.

Cracking Section--

This section contains a single train of equipment to first dry and then thermally decompose the sodium salts from the crystallization section. An oil-fired furnace is provided for heating. A byproduct Na_2SO_4 storage tank with a 7-day storage capacity is provided.

Gypsum Production--

Since makeup Na_2CO_3 is relatively expensive, this section has been included to recover the sodium and, indirectly, to convert the sulfate salt to gypsum. In addition to the equipment to produce gypsum, the equipment needed to pass the byproduct SO_2 from the crackers through the gypsum reactor to form additional $NaHSO_3$ has been included.

Absorbent Regeneration--

Since Na_2SO_3, rather than the $NaHSO_3$ formed in the previous gypsum production section, is the active compound in reducing the absorbed NO_x, the single train of equipment in this section is provided to convert the bisulfite to reusable sulfite in two stages. Initially the solution is heated to strip any excess SO_2 and is then neutralized with limestone slurry to form both Na_2SO_3 and $CaSO_3$. The $CaSO_3$ is separated, reslurried, and sent back to the gypsum production section. The Na_2SO_3 is recycled to the reducing reactor.

Raw Material Storage--

This section includes all the processing equipment needed to receive, store, and retrieve for later use all of the various raw materials (except limestone) needed in the process. Storage capacity is designed for 30 days of normal usage.

IHI SO_x-NO_x PROCESS - WET, OXIDATION-ABSORPTION-REDUCTION

Process Description

The IHI process oxidizes NO to NO_2 by injecting ozone into the flue gas stream (11). This NO_2 is then absorbed into a calcium-based slurry and reacts with the absorbed SO_2 to form a mixture of nitrate salts, molecular nitrogen, and complex nitrogen-sulfur compounds. The complex nitrogen-sulfur compounds are decomposed into molecular nitrogen and calcium sulfate ($CaSO4$). Most of the SO_2 in the flue gas absorbed into solution as sulfite ions are oxidized to sulfates in an oxidizing tower and removed in a centrifuge as byproduct gypsum.

The overall system is very similar to the other wet, oxidation-absorption-reduction processes, in that it consists of six major sections: (1) prescrubbing, (2) gas-phase oxidation, (3) absorption, (4) oxidation of sulfites and reduction of absorbed NO_x, (5) byproduct gypsum recovery, and (6) decomposition of complex nitrogen-sulfur compounds.

The flue gas from the air heater passes through a prescrubber which removes most of the particulates and essentially all of the chlorides. The flue gas is cooled from 300°F to 127°F and humidified by the evaporation of water from the scrubbing solution. Part of the scrubber liquid is purged to a thickener and then to a filter. The filtrate is recycled to the thickener. Most of the thickener overflow is recycled to the prescrubber after a small purge stream has been removed for chloride control. This purge stream passes through a secondary thickener to remove any remaining fly ash and, after being used as wash water for the byproduct gypsum cake, is pumped to the bottom of the absorber.

As the cooled flue gas passes from the prescrubber to the absorber, ozone in a weak ozone-air mixture is injected to selectively oxidize the NO to NO_2 by the following reaction:

$$NO_{(g)} + O_{3(g)} \rightarrow NO_{2(g)} + O_{2(g)} \quad (1)$$

Since an O_3:NO_x mol ratio of only 0.85:1.0 is used for 90% NO_x removal, the ozone injection system must be properly designed to give good mixing characteristics and to prevent the undesirable formation of dinitrogen pentoxide (N_2O_5):

$$2NO_{2(g)} + O_3 \rightarrow N_2O_{5(g)} + O_{2(g)} \qquad (2)$$

Following oxidation, the flue gas is contacted with a limestone slurry in a specially designed countercurrent-flow spray tower. This slurry has a pH of 4-6 and contains low concentrations of copper chloride ($CuCl_2$) and sodium chloride (NaCl) as catalysts to aid in the absorption and reduction of NO_2. The SO_2 is absorbed from the flue gas and undergoes the following liquid-phase reactions:

$$SO_{2(g)} \rightarrow SO_{2(aq)} \qquad (3)$$

$$SO_{2(aq)} + CaCO_{3(s)} + 1/2H_2O \rightarrow CaSO_3\cdot 1/2H_2O_{(s)} + CO_{2(g)} \qquad (4)$$

$$SO_{2(aq)} + CaSO_3\cdot 1/2H_2O_{(s)} + 1/2\ H_2O \rightarrow Ca(HSO_3)_{2(aq)} \qquad (5)$$

The NO_2 is absorbed and undergoes the following liquid-phase reactions:

$$2NO_{2(g)} \rightarrow N_2O_{4(aq)} \qquad (6)$$

$$N_2O_{4(aq)} + 4CaSO_3\cdot 1/2H_2O_{(s)} + 6H_2O \rightarrow N_{2(g)} + 4CaSO_4\cdot 2H_2O_{(s)} \qquad (7)$$

$$\begin{aligned} N_2O_{4(aq)} + Ca(HSO_3)_{2(aq)} + 5CaSO_3\cdot 1/2H_2O_{(s)} + 17/2H_2O \rightarrow \\ Ca(NH_2SO_3)_{2(aq)} + 5CaSO_4\cdot 2H_2O_{(s)} \end{aligned} \qquad (8)$$

$$\begin{aligned} 2N_2O_{4(aq)} + 4CaSO_3\cdot 1/2H_2O_{(s)} \rightarrow \\ Ca(NO_2)_{2(aq)} + Ca(NO_3)_{2(aq)} + 2Ca(HSO_3)_{2(aq)} \end{aligned} \qquad (9)$$

In addition to these primary reactions, the other secondary reactions occur in the absorbing solution. Oxygen absorbed from the flue gas oxidizes sulfite ion by the following reaction:

$$CaSO_3\cdot 1/2H_2O_{(s)} + 1/2O_{2(aq)} + 3/2H_2O \rightarrow CaSO_4\cdot 2H_2O_{(s)} \qquad (10)$$

The NO_x, including some N_2O_5, can also undergo the following reactions:

$$N_2O_{4(aq)} + 3CaSO_3\cdot 1/2H_2O_{(s)} + 3/2H_2O \rightarrow N_2O_{(g)} + 3CaSO_4\cdot 2H_2O_{(s)} \qquad (11)$$

$$N_2O_{5(aq)} + 2CaSO_3\cdot 1/2H_2O_{(s)} \rightarrow Ca(NO_3)_{2(aq)} + Ca(HSO_3)_{2(aq)} \qquad (12)$$

Thus the SO_2, absorbed to form the sulfite ion, is used to reduce the NO_2 to molecular nitrogen and complex nitrogen-sulfur compounds. Approximately 80% of the absorbed NO_x is converted to calcium sulfamates $[Ca(NH_2SO_3)_2]$ and most of the remainder to molecular nitrogen. According to the process developer nitrate salts represent only a small amount (<10%) of the total absorbed NO_x.

After passing through a mist eliminator, the cleaned flue gas is reheated to provide plume buoyany and is then exhausted through the stack.

From the absorber, the absorbent slurry flows into a holding tank where fresh catalyst and makeup limestone slurry are added. Most of the absorbent slurry is recycled to the absorber. A small (1-2% of the circulating slurry) purge stream is acidified with a small amount of H_2SO_4. The acidified slurry is pumped to an oxidizing tower where the $CaSO_3$ in the purge stream is converted to gypsum as it passes countercurrently through an airstream. The gypsum slurry is pumped to the gypsum thickener for initial separation. The thickener bottoms are centrifuged to remove byproduct gypsum and the centrate is recycled to the gypsum thickener.

The overflow from the gypsum thickener, which contains complex nitrogen-sulfur compounds and nitrate salts is neutralized with powdered limestone to precipitate the dissolved catalyst and pumped to a thickener. The bottoms from the thickener, mostly excess limestone and catalyst, are recycled to the absorber. The thickener overflow is pumped to a second neutralization reactor where additional powdered limestone is added to prevent the formation of NH_3. The effluent from the reactor is passed through a double-effect evaporator system to form a mixed salt cake. This cake is passed through a rotary kiln operating at 147-165°F where the sulfamates and the nitrate salts are decomposed to molecular nitrogen and $CaSO_4$. The decomposition of the $Ca(NH_2SO_3)_2$ salts is given by the following reaction:

$$2Ca(NH_2SO_3)_{2(aq)} + 4CaCO_{3(aq)} + 2Ca(HSO_4)_{2(aq)} + 3O_{2(aq)} + 8H_2O \rightarrow 2N_{2(g)} + 8CaSO_4 \cdot 2H_2O_{(s)} + 4CO_{2(g)} \qquad (13)$$

The off-gas from the decomposer, which may contain some SO_2 in addition to the nitrogen and carbon dioxide (CO_2), is recycled to the flue gas ducts.

Analysis of Processing Subsections

The flow diagram and material balance for the IHI process in the base case application are shown in Figure 3 and Table 27 respectively. The IHI SO_x-NO_x system is composed of 10 separate processing sections. The detailed equipment list for the IHI process listed by processing section is shown in Table 28. Each of these subsections is described in more detail in the following subheadings.

Material Handling--

This and the following area, feed preparation, comprise the limestone preparation section. The material-handling section includes all equipment

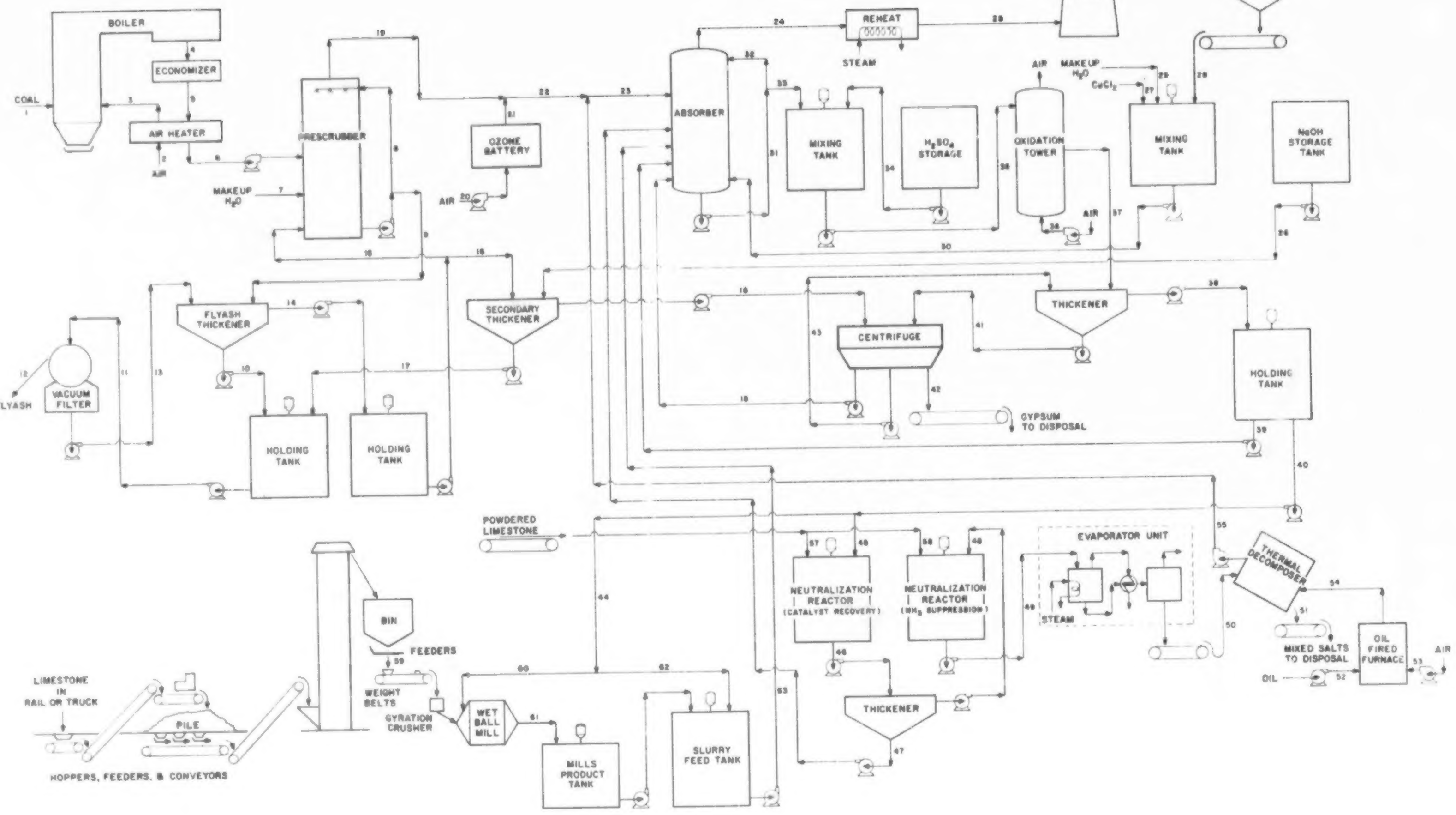
BOILER
COAL
ECONOMIZER
AIR HEATER
AIR
PRESCRUBBER
MAKEUP H_2O
OZONE BATTERY
AIR
ABSORBER
REHEAT
STEAM
STACK
MIXING TANK
H_2SO_4 STORAGE
AIR
OXIDATION TOWER
AIR
MAKEUP H_2O
$CuCl_2$
MIXING TANK
NaCl SILO
NaOH STORAGE TANK
FLYASH
VACUUM FILTER
FLYASH THICKENER
HOLDING TANK
HOLDING TANK
SECONDARY THICKENER
CENTRIFUGE
GYPSUM TO DISPOSAL
THICKENER
HOLDING TANK
POWDERED LIMESTONE
NEUTRALIZATION REACTOR (CATALYST RECOVERY)
NEUTRALIZATION REACTOR (NH_3 SUPPRESSION)
EVAPORATOR UNIT
STEAM
THERMAL DECOMPOSER
MIXED SALTS TO DISPOSAL
OIL
OIL FIRED FURNACE
AIR
THICKENER
LIMESTONE IN RAIL OR TRUCK
PILE
HOPPERS, FEEDERS, & CONVEYORS
BIN
FEEDERS
WEIGHT BELTS
GYRATION CRUSHER
WET BALL MILL
MILLS PRODUCT TANK
SLURRY FEED TANK

Figure 3. IHI SO_x-NO_x process. Flowsheet.

TABLE 27. IHI SO_x-NO_x PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	428,600	4,546,200	4,101,800	4,516,100	4,516,100
2						
3	Sft^3/min (60^{o}F)		1,005,000	906,700	958,000	958,000
4	Temperature, ^{o}F		80	535	890	705
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to prescrubber	Makeup H_2O	Recycle solution to the top of the prescrubber	Prescrubber purge to ash removal	Fly ash thickener underflow
1	Total stream, lb/hr	4,960,400	287,900	N/A	2,860,000	613,800
2						
3	Sft^3/min (60^{o}F)	1,056,000				
4	Temperature, ^{o}F	300		127	127	
5	Pressure, psig					
6	Gpm		575.3			1,160
7	Specific gravity			1.02	1.02	1.06
8	pH			1.5	1.5	
9	Undissolved solids, %			2.0	2.0	8.9
10						

	Stream No.	11	12	13	14	15
	Description	Fly ash slurry feed to filter	Fly ash to disposal	Filtrate recycle to fly ash thickener	Fly ash thickener overflow	Fly ash thickener overflow to prescrubber
1	Total stream, lb/hr	616,000	68,300	547,700	2,794,000	2,728,100
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F	127	127	127	127	
5	Pressure, psig					
6	Gpm	1,161			5,556	
7	Specific gravity	1.06			1.00	
8	pH					
9	Undissolved solids, %	8.9	80	0.0	0.095	
10						

	Stream No.	16	17	18	19	20
	Description	Fly ash thickener overflow to secondary thickener	Secondary thickener underflow to holding tank	Secondary thickener overflow recycled to absorber (via gypsum centrifuge)	Flue gas from prescrubber	Clean air feed to ozone generator
1	Total stream, lb/hr	66,000	2,200	63,800	5,115,700	403,800
2						
3	Sft^3/min (60^{o}F)				1,070,200	88,359
4	Temperature, ^{o}F				127	
5	Pressure, psig					
6	Gpm					
7	Specific gravity		1.02			
8	pH					
9	Undissolved solids, %	0.095	4.1			
10						

(continued)

TABLE 27. (continued)

	Stream No.	21	22	23	24	25
	Description	Air-ozone mixture from generators	Flue gas-ozone mixture	Flue gas to absorber	Flue gas to reheater	Flue gas to stack
1	Total stream, lb/hr	403,800	5,519,632	5,554,396	5,567,314	5,567,314
2						
3	Sft^3/min (60°F)		1,155,805	1,163,085	1,164,625	1,165,790
4	Temperature, °F		127	127	127	175
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	26	27	28	29	30
	Description	Makeup NaOH	Makeup $CuCl_2$	Makeup NaCl	Makeup H_2O	Mixing tank to absorber
1	Total stream, lb/hr	620	22	330	53,600	54,600
2						
3	Sft^3/min (60°F)					
4	Temperature, °F				60	
5	Pressure, psig					
6	Gpm				103	
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	31	32	33	34	35
	Description	Absorber holding tank purge	Recycled absorbent to the top of the absorber	Absorbent solution to regeneration section	Makeup H_2SO_4	Forward feed to oxidation tower
1	Total stream, lb/hr	75,372,000	74,800,000	572,000	220	572,220
2						
3	Sft^3/min (60°F)					
4	Temperature, °F	127	127	127	60	
5	Pressure, psig					
6	Gpm					
7	Specific gravity	1.17	1.17	1.17		
8	pH					
9	Undissolved solids, %	7.93	7.93	7.93	1.84	1.17
10						

	Stream No.	36	37	38	39	40
	Description	Air feed to oxidation tower	Forward feed to gypsum thickener	Gypsum thickener overflow to holding tank	Gypsum thickener overflow recycled to absorber	Gypsum thickener overflow to N_2-treatment and limestone preparation
1	Total stream, lb/hr	15,510	574,400	517,200	246,400	270,800
2						
3	Sft^3/min (60°F)	3,394				
4	Temperature, °F	60	127			
5	Pressure, psig					
6	Gpm					
7	Specific gravity		1.18			
8	pH		3.5			
9	Undissolved solids, %		8.7			
10						

(continued)

TABLE 27. (continued)

Stream No.		41	42	43	44	45
Description		Gypsum thickener underflow	Byproduct gypsum to disposal	Centrate recycled to gypsum thickener	Overflow to limestone slurry feed tank and wet ball mill	Gypsum thickener overflow to No.1 neutralization reactor
1	Total stream, lb/hr	171,600	57,200	114,400	184,000	86,000
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity	1.33				
8	pH					
9	Undissolved solids, %	30	84.3	0.0		
10						

Stream No.		46	47	48	49	50
Description		Effluent from neutralization reactor	Thickener underflow recycled to absorber	Thickener overflow to neutralization reactor No. 2	Evaporator feed	Thermal decomposer feed
1	Total stream, lb/hr	86,100	4,400	81,700	84,900	13,500
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					212
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %				3.8	
10						

Stream No.		51	52	53	54	55
Description		Mixed salts to disposal	Fuel oil to furnace	Combustion air to furnace	Combustion flue gas to decomposer	Off-gas thermal decomposer
1	Total stream, lb/hr	10,200	1,310	27,202	28,510	31,900
2						
3	Sft^3/min (60°F)			5,752	5,971	6,885
4	Temperature, °F	1,562		60	1,562	1,562
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

Stream No.		56	57	58	59	60
Description		Not used	Makeup powdered limestone to neutralization reactor No. 1	Makeup powdered limestone to neutralization reactor No. 2	Makeup limestone from pile	Gypsum thickener overflow to wet ball mill
1	Total stream, lb/hr		140	3,230	33,700	21,700
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

(continued)

TABLE 27. (continued)

Stream No.		61	62	63		
Description		Wet ball mill product to product tank	Gypsum thickener overflow to limestone slurry tank	Fresh makeup limestone slurry to absorber		
1	Total stream, lb/hr	54,400	163,100	217,500		
2						
3	Sft^3/min (60°)					
4	Temperature, $^{\circ}$F					
5	Pressure, psig					
6	Gpm					
7	Specific gravity	1.73		1.23		
8	pH					
9	Undissolved solids, %	60		15		
10						

1						
2						
3						
4						
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6						
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TABLE 28. IHI SO_x-NO_x PROCESS

BASE CASE EQUIPMENT LIST
DESCRIPTION AND COST

Area 1--Material Handling

	Item	No.	Description	Total equipment cost, 1979 $
1.	Car shaker and hoist	1	20 hp shaker, 7.5 hp hoist	31,000
2.	Car puller	1	25 hp puller, 5 hp return	52,000
3.	Hopper, unloading	1	16 ft dia x 10 ft straight side height, carbon steel	12,800
4.	Feeder, unloading vibrating	1	3.5 hp	14,400
5.	Conveyor, unloading belt	1	20 ft horizontal, 5 hp	17,700
6.	Conveyor, unloading incline belt	1	310 ft, 50 hp	89,200
7.	Unloading pit dust collector	1	Polypropylene bag type, 2,200 ft^3/min, 7.5 hp	19,200
8.	Pump, unloading pit sump	1	60 gpm, 70 ft head, 5 hp	4,200
9.	Conveyor, storage belt	1	1,000 ft, 2 hp	45,200
10.	Tripper, storage conveyor	1	30 ft/min, 1 hp	16,400
11.	Mobile equipment	1	Scraper tractor	137,800
12.	Hopper, reclaim	1	7 ft wide x 4.25 ft high, 2 ft wide bottom, carbon steel	1,500
13.	Feeder, reclaim vibrating	1	3.5 hp	14,400
14.	Conveyor, reclaim belt	1	100 ft, 2 hp	32,200

(continued)

TABLE 28. (continued)

Item	No.	Description	Total equipment cost, 1979 $
15. Conveyor, reclaim incline belt	1	193 ft, 40 hp	53,800
16. Dust collector, reclaim pit	1	Polypropylene bag type	19,200
17. Pump, reclaimer pit sump	1	60 gpm, 70 ft head, 5 hp	4,200
18. Elevator, reclaim bucket	1	90 ft high, 75 hp	86,400
19. Bin, feed	2	13 ft dia x 21 ft straight side height, covered, carbon steel	33,200
Subtotal			684,800

Area 2--Feed Preparation

Item	No.	Description	Total equipment cost, 1979 $
1. Feeder, weigh bin	2	14 ft pulley centers, 2 hp	39,600
2. Gyratory crushers	2	75 hp	119,500
3. Ball mill dust collectors	2	Polypropylene bag type, 2,200 ft^3/min, 7.5 hp	38,400
4. Ball mill	2	9.3 tons/hr, 125 hp	320,100
5. Tank, ball mill product	2	10 ft dia x 10 ft high, 5,500 gal, open top, flakeglass-lined, carbon steel	26,800
6. Agitator, mills product tank	2	10 hp	17,500
7. Pump, ball mill product tank	3	Centrifugal belt drive, 35 gpm, 50 ft head, 1.5 hp, resin-lined, carbon steel	7,100

(continued)

TABLE 28. (continued)

Item	No.	Description	Total equipment cost, 1979 $
8. Tank, limestone slurry feed	1	31 ft dia x 32 ft high, 180,700 gal, open top, resin-lined, carbon steel	100,900
9. Pump, limestone slurry feed	2	Centrifugal belt drive, 355 gpm, 100 ft head, 20 hp, resin-lined, carbon steel	11,500
10. Agitator, limestone slurry tank	1	60 hp, resin-lined, carbon steel	58,700
Subtotal			740,100

Area 3--Gas Handling

Item	No.	Description	Total equipment cost, 1979 $
1. Blower, flue gas	2	772,000 aft^3/min, 4,000 hp, ΔP 21 in. H_2O, 316 SS	1,596,300
Subtotal			1,596,300

Area 4--Particulate Control

Item	No.	Description	Total equipment cost, 1979 $
1. Prescrubber	1	Venturi with spray, 64.9 ft dia x 105 ft high, resin-lined, carbon steel	4,788,600
2. Pump, prescrubber recycle	4	Centrifugal belt drive, 13,700 gpm, 75 ft head, 500 hp, resin-lined, carbon steel	187,900

(continued)

TABLE 28. (continued)

Item	No.	Description	Total equipment cost, 1979 $
3. Thickener, primary fly ash	1	98.4 ft dia x 11.6 ft high, 88,200 gal, resin-lined, carbon steel	650,500
4. Pump, primary thickener underflow	2	Centrifugal belt drive, 1,160 gpm, 75 ft head, 40 hp, resin-lined, carbon steel	19,300
5. Tank, filter feed	1	35 ft dia x 40 ft high, 287,900 gal, open top, resin-lined, carbon steel	60,200
6. Agitator, holding tank	1	50 hp, resin-lined, carbon steel	51,100
7. Pump, filter feed	2	Centrifugal belt drive, 1,165 gpm, 50 ft head, 30 hp, resin-lined, carbon steel	16,300
8. Ash filter	5	Vacuum, 485 ft^2 filtering area	819,000
9. Pump, filtrate recycle	2	Centrifugal belt drive, 1,090 gpm, 50 ft head, 25 hp, resin-lined, carbon steel	15,800
10. Pump, thickener overflow to holding tank	1	Centrifugal belt drive, 5,600 gpm, 75 ft head, 200 hp, resin-lined, carbon steel	68,900
11. Tank, primary thickener overflow holding	1	48 ft dia x 50 ft high, 676,800 gal, open top, resin-lined, carbon steel	243,500
12. Agitator, holding tank	1	35 hp, resin-lined, carbon steel	39,900
13. Pump, secondary thickener feed	2	Centrifugal belt drive, 5,600 gpm, 75 ft head, 200 hp, resin-lined, carbon steel	68,900

(continued)

TABLE 28. (continued)

Item	No.	Description	Total equipment cost, 1979 $
14. Thickener, secondary fly ash	1	21.3 ft dia x 9.88 ft high, 26,200 gal, resin-lined, carbon steel	100,500
15. Pump, secondary thickener underflow	2	Centrifugal belt drive, 5 gpm, 75 ft head, 0.25 hp, resin-lined, carbon steel	4,600
16. Pump, secondary thickener overflow	2	Centrifugal belt drive, 130 gpm, 50 ft head, 3 hp motor, resin-lined, carbon steel	6,700
17. Conveyor, fly ash to reslurrying tank	1	Belt, 14 in. wide x 100 ft long, 7.5 hp, 30 ft lift, 34.1 tons/hr	25,500
18. Tank reslurrying	1	22 ft dia x 24 ft high, 68,200 gal, open top, resin-lined, carbon steel	28,000
19. Agitator, reslurrying tank	1	20 hp, resin-lined, carbon steel	27,700
20. Pump, fly ash slurry	4	Centrifugal belt drive, 665 gpm, 100 ft head, 40 hp, resin-lined, carbon steel	26,800
21. Pump, pond return	3	Centrifugal belt drive, 600 gpm, 100 ft head, 25 hp, resin-lined, carbon steel	19,700
Subtotal			7,269,400

Area 5--SO_x-NO_x Absorption

Item	No.	Description	Total equipment cost, 1979 $
1. Absorber	1	Spray tower, 53.5 ft dia x 139.4 ft high, resin-lined, carbon steel	4,268,300

(continued)

TABLE 28. (continued)

	Item	No.	Description	Total equipment cost, 1979 $
2.	Pump, absorber recycle	9	Centrifugal belt drive, 21,500 gpm, 75 ft head, 800 hp, resin-lined, carbon steel	966,100
	Subtotal			5,234,400

Area 6--Reheat

	Item	No.	Description	Total equipment cost, 1979 $
1.	Reheater	1	Indirect steam, 20,000 ft^2, 316 SS	636,200
2.	Soot blowers	20	Air, fixed	130,000
	Subtotal			766,200

Area 7--Gypsum Production Section

	Item	No.	Description	Total equipment cost, 1979 $
1.	Tank, H_2SO_4 storage	1	12 ft dia x 15 ft high, 12,700 gal, closed top, carbon steel	11,100
2.	Pump, H_2SO_4 feed	2	Centrifugal belt drive, 0.25 gpm, 50 ft head, 0.25 hp, carbon steel	4,500
3.	Tank, slurry-acid mixing	1	22 ft dia x 24 ft high, 68,100 gal, open top, resin-lined, carbon steel	53,200
4.	Agitator, slurry-acid mix tank	1	25 hp, resin-lined, carbon steel	31,000

(continued)

TABLE 28. (continued)

Item	No.	Description	Total equipment cost, 1979 $
5. Pump, oxidation tower feed	2	Centrifugal belt drive, 980 gpm, 75 ft head, 40 hp, resin-lined, carbon steel	14,500
6. Blower, air feed to oxidation tower	1	3,400 aft^3/min, 12.5 hp, ΔP 15 in. H_2O, carbon steel	5,300
7. Oxidation tower	1	Self-supporting, vertical tower with smoke atomizers, 19.7 ft dia x 42.6 ft high, flakeglass-lined, carbon steel	67,100
8. Pump, gypsum thickener feed	2	Centrifugal belt drive, 975 gpm, 50 ft head, 25 hp, resin-lined, carbon steel	13,000
9. Thickener, gypsum	1	50.8 ft dia x 10.4 ft high, 157,100 gal, resin-lined, carbon steel	274,100
10. Pump, gypsum thickener underflow	2	Centrifugal belt drive, 260 gpm, 50 ft head, 7.5 hp, resin-lined, carbon steel	10,300
11. Separator, gypsum	21	Centrifuge	1,127,700
12. Pump, centrate recycle	2	Centrifugal belt drive, 205 gpm, 50 ft head, 5 hp, resin-lined, carbon steel	7,000
13. Pump, wash solution recycle	2	Centrifugal belt drive, 130 gpm, 75 ft head, 4 hp, resin-lined, carbon steel	7,000
14. Conveyor, gypsum to outdoor storage	1	Belt, 14 in. wide x 300 ft long, 10 hp, 50 ft lift, 28.6 tons/hr	11,400
15. Pump, gypsum thickener overflow	2	Centrifugal belt drive, 925 gpm, 100 ft head, 50 hp, resin-lined, carbon steel	15,300
Subtotal			1,652,500

(continued)

TABLE 28. (continued)

Area 8--Nitrate Treatment

Item	No.	Description	Total equipment cost, 1979 $
1. Reactor, neutralization catalyst recovery	1	22.5 ft dia x 25 ft high, 23,700 gal, open top, resin-lined, carbon steel	58,900
2. Agitator, neutralization reactor No. 1	1	20 hp, resin-lined, carbon steel	27,700
3. Pump, effluent feed to thickener	2	Centrifugal belt drive, 155 gpm, 50 ft head, 4 hp, resin-lined, carbon steel	7,000
4. Thickener, catalyst recovery	1	21.3 ft dia x 34.7 ft high, 92,800 gal, resin-lined, carbon steel	167,600
5. Pump, thickener underflow recycle	2	Centrifugal belt drive, 10 gpm, 75 ft head, 0.5 hp, resin-lined, carbon steel	4,600
6. Pump, thickener overflow	2	Centrifugal belt drive, 150 gpm, 50 ft head, 4 hp, resin-lined, carbon steel	7,000
7. Reactor, neutralization reactor NH_3 suppression	1	22 ft dia x 25 ft high, 22,600 gal, open top, resin-lined, carbon steel	55,100
8. Agitator, neutralization reactor No. 2	1	20 hp, resin-lined, carbon steel	27,700
9. Pump, effluent feed to evaporator No. 1	2	Centrifugal belt drive, 150 gpm, 50 ft head, 4 hp, resin-lined, carbon steel	7,000
10. Evaporator	7	Double effect, 7.5 ft dia x 8 ft high, 2,600 gal, 316 SS	1,985,200
11. Conveyor, evaporator to decomposer	1	Belt, 14 in. wide x 100 ft long, 1 hp, 30 ft lift, 7.0 tons/hr	25,100

(continued)

TABLE 28. (continued)

Item	No.	Description	Total equipment cost, 1979 $
12. Thermal decomposer	1	Rotary kiln, direct heat, 5.9 ft dia x 82 ft long, 316 SS	460,000
13. Blower, decomposer off-gas	1	26,800 aft^3/min, 100 hp, ΔP 14 in. H_2O, 316 SS	17,600
14. Conveyor, mixed salts to temporary storage	1	Screw, 14 in. x 40 ft long, 15 hp, 10 tons/hr	5,100
15. Conveyor, mixed salts to disposal	1	Belt, 14 in. wide x 300 ft long, 1.5 hp, 50 ft lift, 5.1 tons/hr	79,300
16. Tank, fuel oil storage	1	27.5 ft dia x 28 ft high, 124,200 gal, closed top, carbon steel	93,900
17. Pump, fuel oil	2	Centrifugal belt drive, 3 gpm, 100 ft head, 0.25 hp	4,600
18. Oil-fired furnace	1	27 MBtu/hr heat load	167,000
Subtotal			3,200,400

Area 9--Raw Material Storage

Item	No.	Description	Total equipment cost, 1979 $
1. Tank, NaOH storage	1	15 ft dia x 20 ft high, 26,400 gal, closed top, resin-lined, carbon steel	19,100
2. Pump, NaOH feed	2	Centrifugal belt drive, 0.6 gpm, 50 ft head, 0.26 hp, resin-lined, carbon steel	4,600
3. Tank, mixing	1	20 ft dia x 22 ft high, 51,700 gal, open top, resin-lined, carbon steel	44,600

(continued)

TABLE 28. (continued)

Item	No.	Description	Total equipment cost, 1979 $
4. Agitator, mixing tank	1	15 hp, resin-lined, carbon steel	21,600
5. Pump, additive solution feed	2	Centrifugal belt drive, 110 gpm, 50 ft head, 2.5 hp, resin-lined, carbon steel	107,400
6. Conveyor, feed to limestone silo	1	Bucket, 6 in. x 4 in. x 4-1/4 in., 3 hp, 75 ft lift, 2.0 tons/hr	10,300
7. Silo, powdered limestone	1	Conical bottom, 20 ft dia x 40 ft high, reinforced concrete	66,000
8. Weigh feeder, powdered limestone	1	3,300 lb/hr	6,300
9. Conveyor, limestone to neutralization reactor No. 2	1	Belt, 14 in. wide x 100 ft long, 0.5 hp, 25 ft lift, 1.6 tons/hr	24,800
Subtotal			304,700

Area 10--Ozone Generating System

Item	No.	Description	Total equipment cost, 1979 $
1. Ozone generation package	1	Includes all equipment except instrumentation and controls	76,984,000
Subtotal			76,984,000
Total, Areas 1-10			98,432,800

needed to receive and unload limestone from rail or truck transport and the equipment needed to supply limestone to the weigh feeders. Only one train of processing equipment is used.

Feed Preparation--

The feed preparation area consists of one train of equipment to weigh, crush, and slurry the makeup limestone. The fresh limestone slurry, containing 15 weight percent solids, is stored in the limestone slurry feed tank with a surge volume sufficient for 8 hours of normal operation.

Gas Handling--

Two ID fans are provided between the reheating systems and the stack to overcome the pressure drop through the IHI SO_x-NO_x process. Each fan has been sized to handle one-half of the total flue gas volume at a pressure drop of 21 inches water.

Particulate Control--

A single venturi-spray tower is used. The venturi is designed with sufficient liquid surge volume to make a prescrubber holding tank unnecessary. A single train fly ash slurry thickener and filter system is used.

SO_x-NO_x Absorption--

Simultaneous SO_x-NO_x absorption is carried out in a single spray tower, designed with sufficient liquid retention time in the base of the tower to make a separate absorbent holding tank unnecessary.

Reheat--

Flue gas reheat to 175°F is provided by a single reheater using an indirect steam system.

Gypsum Production--

During normal operation most of the SO_2 is converted to $CaSO_3$ and only some is oxidized to gypsum. Since this mixture will not form a solid material without further treatment, the purge from the absorber is air-oxidized to convert the sulfite-sulfate mixture to sulfate.

Nitrogen Treatment Section--

Since significant amounts of the absorbed NO_x are converted to both nitrate salts and complex nitrogen-sulfur compounds, a purge stream containing these mixed salts is neutralized to remove the dissolved catalyst, evaporated, and the resulting mixed salts thermally decomposed. The elevated temperatures for decomposing these mixed salts are obtained by using the flue gas from an oil-fired furnace. This purge stream is relatively small and the equipment is designed so that only one train of equipment is required.

Raw Material Storage--

This section includes all the processing equipment needed to receive, store, and retrieve for later use all of the various raw materials (except limestone) needed in the process. Storage capacity is designed for 30 days of normal usage.

Oxone Generation Section--

The ozone system is designed as a package unit to include all of the process equipment needed to generate ozone from air (13).

MORETANA CALCIUM SO_x-NO_x PROCESS - WET, OXIDATION-ABSORPTION-REDUCTION

Process Description

The Moretana Calcium process is a wet, oxidation-absorption-reduction system using ClO_2 for the gas-phase oxidation of NO to NO_2 (1). Although the use of ClO_2 results in the formation of undesirable nitrate and chloride salts in the scrubbing solution, it has the advantages of costing less to produce and of requiring a lower oxidant to NO_x mol ratio. The overall process consists of five basic steps: (1) prescrubbing, (2) gas-phase oxidation, (3) simultaneous SO_x-NO_x absorption, (4) byproduct sludge removal, and (5) byproduct nitrate removal.

The flue gas from the air heater initially passes through an ESP which removes 94% of the fly ash. Further fly ash removal as well as chloride removal and humidification is achieved by passing the flue gas through a prescrubber countercurrent to an aqueous scrubbing solution. The liquid effluent from the prescrubber is recirculated after makeup water has been added. A small purge stream of scrubbing liquid is pumped to a pH-controlling tank where a slaked lime slurry is used to neutralize both the prescrubber purge and a waste stream from the ClO_2 plant. The resulting effluent from the pH tank is sent to the fly ash pond.

The flue gas, after passing through a mist eliminator, is injected with a ClO_2-air mixture. The ClO_2 is generated onsite and is injected at a rate of 10% in excess of the stoichiometric requirement. With the proper design of the injection system, the ClO_2 selectively oxidizes the NO by the following reaction:

$$2NO_{(g)} + ClO_{2(g)} + H_2O_{(g)} \rightarrow NO_{2(g)} + HNO_{3(g)} + HCl_{(g)} \qquad (1)$$

This gas-phase reaction occurs very rapidly and completely oxidizes the NO to NO_2 and HNO_3.

The oxidized flue gas is then passed countercurrently to a limestone slurry in a Moretana plate tower absorber. The limestone slurry has a pH of approximately 5.5. It is a mixture of calcium salts, mainly $CaSO_3$, and contains a catalyst to aid in the reduction of NO_x. As the slurry passes through the absorber, the SO_2 is stripped from the flue gas and the following reactions occur in the scrubbing solution:

$$SO_{2(g)} \rightarrow SO_{2(aq)} \qquad (2)$$

$$CaCO_{3(g)} + SO_{2(aq)} \rightarrow CaSO_{3(aq)} + CO_{2(g)} \qquad (3)$$

$$CaSO_{3(aq)} + SO_{2(aq)} + H_2O \rightarrow Ca(HSO_3)_{2(aq)} \qquad (4)$$

At the same time the gases formed from the oxidation of NO are also absorbed from the flue gas stream and undergo the following hypothesized reactions:

$$2NO_{2(g)} \rightarrow N_2O_{4(g)} \tag{5}$$

$$N_2O_{4(g)} \rightarrow N_2O_{4(aq)} \tag{6}$$

$$2N_2O_{4(aq)} + 2CaSO_{3(aq)} + 2CaCO_{3(s)} + 2H_2O \xrightarrow{catalyst} 2Ca(NO_2)_{2(aq)} + 2CaSO_4 \cdot 2H_2O_{(s)} + 2CO_{2(g)} \tag{7}$$

$$Ca(HSO_3)_{2(aq)} + Ca(NO_2)_{2(aq)} + H_2O \xrightarrow{catalyst} N_{2(g)} + 2CaSO_4 \cdot 2H_2O_{(s)} \tag{8}$$

$$HNO_{3(g)} \rightarrow HNO_{3(aq)} \tag{9}$$

$$CaCO_{3(s)} + 2HNO_{3(aq)} \rightarrow Ca(NO_3)_{2(aq)} + H_2O + CO_{2(g)} \tag{10}$$

$$HCl_{(g)} \rightarrow HCl_{(aq)} \tag{11}$$

$$2HCl_{(aq)} + CaCO_{3(s)} \rightarrow CaCl_{2(aq)} + CO_{2(g)} + H_2O \tag{12}$$

Secondary reactions also occur in the absorber; for example, oxygen from the flue gas is absorbed into the scrubbing solution and causes the oxidation of $CaSO_3$ to gypsum:

$$CaSO_{3(aq)} + 1/2O_{2(aq)} + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O_{(s)} \tag{13}$$

However, this oxidation is not as serious a problem in the calcium system as in the sodium systems because $CaSO_3$ is relatively insoluble in the scrubbing solution. Hence, most of the $CaSO_3$ is not present in the solution in a readily oxidizable form. After passing through a mist eliminator, the cleaned flue gas is reheated and vented to the stack.

The liquid effluent from the absorber is recirculated through the absorber after fresh makeup limestone slurry has been added. A small purge stream of the slurry is pumped to a centrifuge where the relatively insoluble $CaSO_3$ and $CaSO_4$ are separated from the soluble calcium chloride ($CaCl_2$) and $Ca(NO_3)_2$. The solids are reslurried and pumped to a sludge pond. The centrate containing the soluble calcium salts is separated into two streams, the largest of which is pumped to the limestone preparation section to transport makeup limestone and makeup catalyst to the absorber holding tank. The second stream is purged to a byproduct nitrate treatment section.

The mixed solution of calcium salts is treated by reacting the solution with $(NH_4)_2SO_4$ to produce a liquid fertilizer by the following reactions:

$$CaCl_{2(aq)} + (NH_4)_2SO_{4(aq)} + 2H_2O \rightarrow 2NH_4Cl_{(aq)} + CaSO_4 \cdot 2H_2O_{(s)} \qquad (14)$$

$$Ca(NO_3)_{2(aq)} + (NH_4)_2SO_{4(aq)} + 2H_2O \rightarrow 2NH_4NO_{3(aq)} + CaSO_4 \cdot 2H_2O_{(s)} \qquad (15)$$

The effluent gypsum slurry is pumped to a centrifuge and the resulting solids are sent to the sludge pond. The centrate containing the dissolved ammonium salts enters a double-effect evaporator system where the salt concentration is increased from 15 to about 50 weight percent. This concentrated solution can be stored and used as a fertilizer material.

Analysis of Processing Subsections

The flow diagram and material balance for the Moretana Calcium process in the base case application are shown in Figure 4 and Table 29 respectively. The Moretana Calcium process is a wet simultaneous SO_x-NO_x system, which is composed of 11 separate processing subsections. As an initial step, the various pieces of process equipment were apportioned into the appropriate subsection. Using the information available from the material balance, each piece of equipment was described, sized, costed, and then listed by processing subsection. This detailed equipment list for the Moretana Calcium process is shown in Table 30. Each of these subsections is described in more detail in the following sections.

Material Handling--

This and the following area, feed preparation, comprise the limestone preparation section. The material-handling portion includes all of the equipment needed to receive and unload the limestone which arrives by either rail or truck and also the equipment needed to maintain a readily available supply of limestone to the weigh feeders. The equipment is sized such that only one train of processing equipment is required.

Feed Preparation--

The feed preparation area consists of a single train of equipment required to weigh, crush, and slurry the makeup limestone. The fresh limestone slurry, containing 8% solids (by weight), is temporarily stored in the limestone slurry feed tank which has a surge volume sufficient for 8 hours of normal operation.

Gas Handling--

Two forced-draft (FD) fans are provided between the ESP and the prescrubber to overcome the pressure drop through the Moretana Calcium process. Each fan has been sized to handle one-half of the total flue gas volume and a pressure drop of 18.5 inches water.

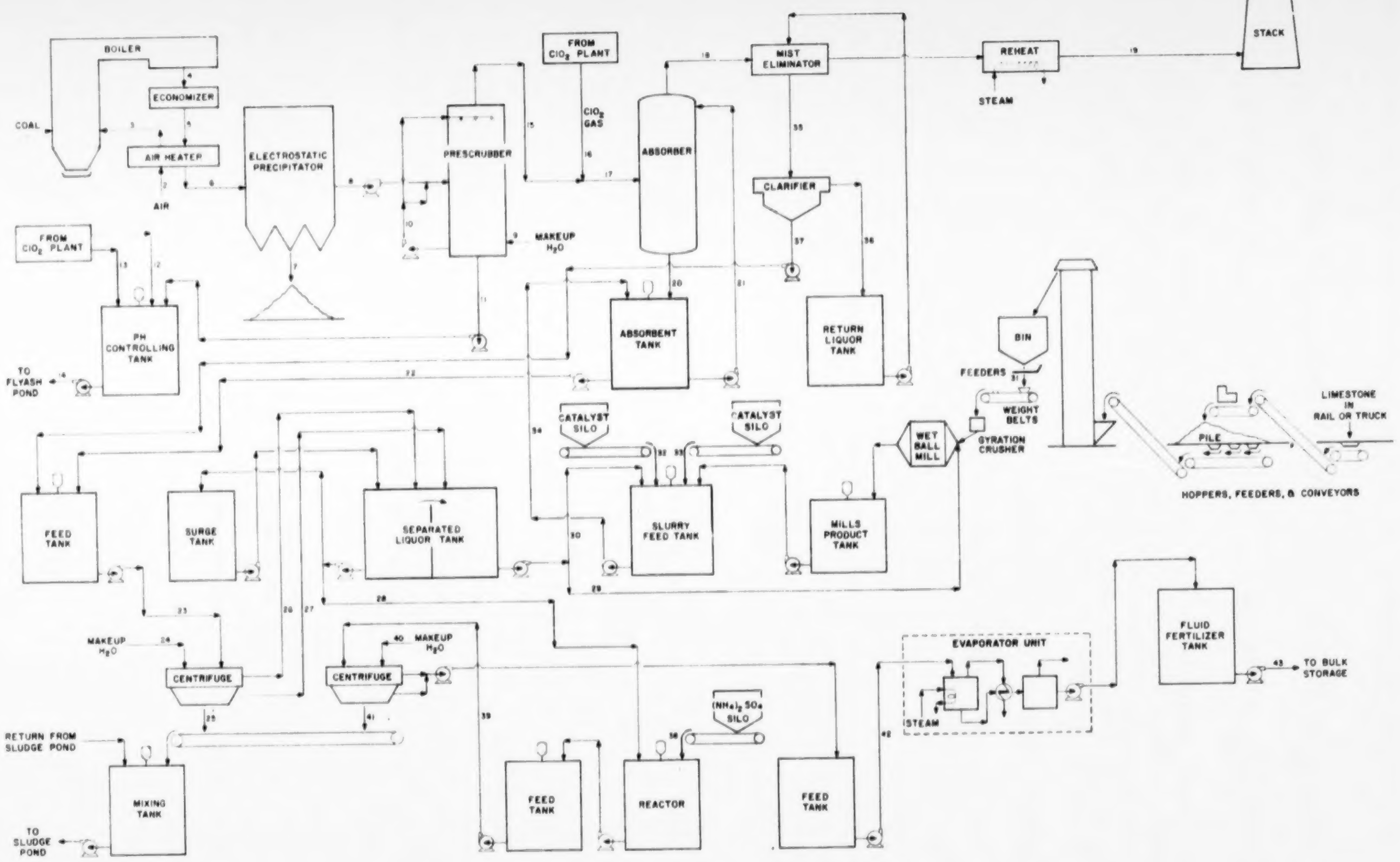
BOILER
COAL
ECONOMIZER
AIR HEATER
AIR
ELECTROSTATIC PRECIPITATOR
PRESCRUBBER
MAKEUP H_2O
FROM ClO_2 PLANT
ClO_2 GAS
ABSORBER
MIST ELIMINATOR
REHEAT
STEAM
STACK
CLARIFIER
RETURN LIQUOR TANK
ABSORBENT TANK
FROM ClO_2 PLANT
PH CONTROLLING TANK
TO FLYASH POND
BIN
FEEDERS
WEIGHT BELTS
GYRATION CRUSHER
WET BALL MILL
LIMESTONE IN RAIL OR TRUCK
PILE
HOPPERS, FEEDERS, & CONVEYORS
CATALYST SILO
CATALYST SILO
SLURRY FEED TANK
MILLS PRODUCT TANK
FEED TANK
SURGE TANK
SEPARATED LIQUOR TANK
MAKEUP H_2O
CENTRIFUGE
MAKEUP H_2O
CENTRIFUGE
RETURN FROM SLUDGE POND
MIXING TANK
TO SLUDGE POND
FEED TANK
REACTOR
$(NH_4)_2SO_4$ SILO
FEED TANK
EVAPORATOR UNIT
STEAM
FLUID FERTILIZER TANK
TO BULK STORAGE

Figure 4. Moretana Calcium process. Flowsheet.

TABLE 29. MORETANA CALCIUM SO_x-NO_x PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	428,600	4,546,200	4,101,800	4,516,100	4,516,100
2						
3	$Sft^3/min\ (60^oF)$		1,005,000	906,700	958,000	958,000
4	Temperature, ^oF		80	535	890	705
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to ESP	Fly ash from ESP	Gas to prescrubber	Makeup H_2O to prescrubber	Recycle slurry to prescrubber
1	Total stream, lb/hr	4,960,400	51,600	4,908,800	332,300	20,429,200
2						
3	$Sft^3/min\ (60^oF)$	1,056,000				
4	Temperature, ^oF	300				
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	11	12	13	14	15
	Description	Prescrubber slurry purged to ash pond	Waste stream from ClO_2 generator	Slaked lime for pH control	Neutralized effluent to fly ash pond	Effluent flue gas from the prescrubber
1	Total stream, lb/hr	61,000	6,370	7,840	125,400	5,115,800
2						
3	$Sft^3/min\ (60^oF)$					1,079,819
4	Temperature, ^oF				127	127
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %				43.75	
10						

	Stream No.	16	17	18	19	20
	Description	ClO_2 gas to flue gas ducts	Oxidized flue gas to absorber	Effluent flue gas from absorber	Reheated flue gas to the stack	Spent absorbent from absorber
1	Total stream, lb/hr	53,300	5,169,000	5,138,900	5,133,700	64,661,200
2						
3	$Sft^3/min\ (60^oF)$	11,352	1,093,129	1,084,700	1,084,700	
4	Temperature, ^oF	85	127	127	175	
5	Pressure, psig					
6	Gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

(continued)

TABLE 29. (continued)

	Stream No.	21	22	23	24	25
	Description	Recycle absorbent to absorber	Purge to regeneration section	Forward feed to gypsum centrifuge	Wash H_2O to gypsum centrifuge	Sulfite-sulfate sludge to disposal pond
1	Total stream, lb/hr	64,680,000	554,300	586,700	75,900	92,400
2						
3	Sft^3/min (60°F)					
4	Temperature, °F	127	127	127		60
5	Pressure, psig					
6	Gpm					
7	Specific gravity	1.16	1.16	1.16	1.0	1.2
8	pH	5.0	5.5	5.5	7	5.5
9	Undissolved solids, %					70
10						

	Stream No.	26	27	28	29	30
	Description	Centrate to separated liquor tank	Wash H_2O to separated liquor tank	Purge to nitrate treatment section	Waste absorbent to wet ball mill	Waste absorbent to slurry feed tank
1	Total stream, lb/hr	471,100	99,100	48,700	32,700	488,800
2						
3	Sft^3/min (60°F)					
4	Temperature, °F	127	86	127		127
5	Pressure, psig					
6	Gpm					
7	Specific gravity	1.10	1.03	1.10	127	1.09
8	pH	5.5	5.5	5.5	5.5	5.5
9	Undissolved solids, %					
10						

	Stream No.	31	32	33	34	35
	Description	Makeup limestone from pile	Makeup catalyst to makeup slurry tank	Makeup catalyst to makeup slurry tank	Fresh makeup slurry to absorbent tank	Effluent slurry from the mist eliminator
1	Total stream, lb/hr	49,000	680	16	571,200	296,300
2						
3	Sft^3/min (60°F)					
4	Temperature, °F	60	60	60	104-122	127
5	Pressure, psig					
6	Gpm					
7	Specific gravity	2.9	0.6-0.9	1.2-1.3	1.15	1.10
8	pH				6	5
9	Undissolved solids, %				8	
10						

	Stream No.	36	37	38	39	40
	Description	Clarifier overflow to return liquor tank	Clarifier underflow to gypsum feed tank	Makeup $(NH_4)_2SO_4$ to gypsum production reactor	Effluent to gyspum centrifuge	Wash H_2O to gypsum centrifuge
1	Total stream, lb/hr	264,000	32,300	8,100	56,800	5,000
2						
3	Sft^3/min (60°F)					
4	Temperature, °F	127	127	60	127	60
5	Pressure, psig					
6	Gpm					
7	Specific gravity	1.10	1.16	1.76	1.17	1.0
8	pH	5	5.5		7	7
9	Undissolved solids, %					
10						

(continued)

TABLE 29. (continued)

Stream No.		41	42	43		
	Description	Gypsum to sludge pond	Centrate to evaporators	Fluid fertilizer to bulk storage		
1	Total stream, lb/hr	11,200	50,600	16,000		
2						
3	Sft^3/min (60°F)					
4	Temperature, °F	60	122	158		
5	Pressure, psig					
6	Gpm	20.3				
7	Specific gravity	1.1	1.06	1.24		
8	pH	7	7	7		
9	Undissolved solids, %					
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

TABLE 30. MORETANA CALCIUM PROCESS

BASE CASE EQUIPMENT LIST
DESCRIPTION AND COST

Area 1--Material Handling

Item	No.	Description	Total equipment cost, 1979 $
1. Car shaker and hoist	1	20 hp shaker, 7.5 hp hoist	31,000
2. Car puller	1	25 hp puller, 5 hp return	52,000
3. Hopper, unloading	1	16 ft dia x 10 ft straight side height, carbon steel	12,800
4. Feeder, unloading vibrating	1	3.5 hp	14,400
5. Conveyor, unloading belt	1	20 ft horizontal, 5 hp	17,700
6. Conveyor, unloading incline belt	1	310 ft, 50 hp	89,200
7. Dust collector, unloading pit	1	Polypropylene bag type, 2,200 ft^3/min, 7.5 hp	19,200
8. Pump, unloading pit sump	1	60 gpm, 70 ft head, 5 hp	4,200
9. Conveyor, storage belt	1	200 ft, 5 hp	76,800
10. Tripper, storage conveyor	1	30 ft^3/min, 1 hp	16,400
11. Mobile equipment	1	Scraper tractor	137,800
12. Hopper, reclaim	2	7 ft wide x 4.25 ft high, 2 ft wide bottom, carbon steel	3,000
13. Feeder, reclaim vibrating	2	3.5 hp	28,700

(continued)

TABLE 30. (continued)

Item	No.	Description	Total equipment cost, 1979 $
14. Conveyor, reclaim belt	1	200 ft, 5 hp	50,700
15. Conveyor, reclaim incline belt	1	193 ft, 40 hp	53,800
16. Dust collector, reclaim pit	1	Polypropylene bag type	19,200
17. Pump, reclaim pit sump	1	60 gpm, 70 ft head, 5 hp	4,200
18. Bucket elevator, reclaim	1	90 ft high, 75 hp	86,400
19. Bin, feed	2	13 ft dia x 21 ft straight side height, covered, carbon steel	33,200
Subtotal			750,700

Area 2--Feed Preparation

Item	No.	Description	Total equipment cost, 1979 $
1. Feeder, bin weigh	2	14 ft pulley centers, 2 hp	39,600
2. Gyratory crushers	2	75 hp	119,500
3. Dust collectors, ball mill	2	Polypropylene bag type, 2,200 ft^3/min, 7.5 hp	38,400
4. Ball mill	2	12.7 tons/hr, 172 hp	358,500
5. Tank, ball mill product	2	10 ft dia x 10 ft high, 5,500 gal, open top, flakeglass-lined, carbon steel	26,800

(continued)

TABLE 30. (continued)

	Item	No.	Description	Total equipment cost, 1979 $
6.	Agitator, ball mill product tank	2	10 hp, resin-lined, carbon steel	30,600
7.	Pump, ball mill product tank	3	Centrifugal belt drive, 110 gpm, 50 ft head, 5 hp, resin-lined, carbon steel	10,400
8.	Tank, limestone slurry feed	1	42.5 ft dia x 45 ft high, 477,500 gal, open top, resin-lined, carbon steel	194,100
9.	Agitator, limestone slurry tank	1	75 hp, resin-lined, carbon steel	68,400
10.	Pump, limestone slurry feed	2	Centrifugal belt drive, 1,000 gpm, 100 ft head, 50 hp, resin-lined, carbon steel	15,300
	Subtotal			901,600

Area 3--Gas Handling

	Item	No.	Description	Total equipment cost, 1979 $
1.	Blower, flue gas	2	Double suction, turbofan, 3,500 hp, ΔP 18.5 in. H_2O, carbon steel	1,472,700
	Subtotal			1,472,700

Area 4--Particulate Control

	Item	No.	Description	Total equipment cost, 1979 $
1.	Prescrubber	2	Moretana tower	893,400

(continued)

TABLE 30. (continued)

Item	No.	Description	Total equipment cost, 1979 $
2. Pump, prescrubber recycle	3	Centrifugal belt drive, 9,950 gpm, 75 ft head, 400 hp, resin-lined, carbon steel	150,900
3. Pump, prescrubber purge	2	Centrifugal belt drive, 60 gpm, 75 ft head, 2 hp, resin-lined, carbon steel	6,500
4. Tank, pH controlling	1	13 ft dia x 15 ft high, 14,900 gal, open top, flakeglass-lined, carbon steel	9,800
5. Agitator, pH tank	1	7.5 hp, resin-lined, carbon steel	13,000
6. Pump, pH tank purge	2	Centrifugal belt drive, 150 gpm, 300 ft head, 20 hp, resin-lined, carbon steel	17,800
Subtotal			1,091,400

Area 5--SO_x-NO_x Absorption

Item	No.	Description	Total equipment cost, 1979 $
1. Absorber	2	Moretana tower	1,118,300
2. Tank absorbent	2	45 ft dia x 50 ft high, 594,800 gal, flakeglass-lined, concrete	451,000
3. Pump, absorber recycle	3	Centrifugal belt drive, 14,000 gpm, 85.3 ft head, 700 hp, resin-lined, carbon steel	155,400

(continued)

TABLE 30. (continued)

Item	No.	Description	Total equipment cost, 1979 $
4. Pump, absorbent tank purge	3	Centrifugal belt drive, 500 gpm, 50 ft head, 15 hp, resin-lined, carbon steel	16,700
5. Agitator, absorbent tank	2	60 hp, resin-lined, carbon steel	120,100
6. Mist eliminator	2	Euroform T-102-2, polypropylene with flakeglass-lined, carbon steel housing, 26.2 ft x 26.2 ft x 26.2 ft	286,100
7. Clarifier, with rake and drive	2	45 ft dia x 10 ft high, 119,000 gal, flakeglass-lined, carbon steel	529,700
8. Pump, clarifier overflow	3	Centrifugal belt drive, 30 gpm, 50 ft head, 1 hp, resin-lined, carbon steel	7,000
9. Tank, return liquor	2	17 ft dia x 19 ft high, 14,900 gal, open top, flakeglass-lined, carbon steel	63,400
10. Pump, return liquor	3	Centrifugal belt drive, 250 gpm, 50 ft head, 7.5 hp, resin-lined, carbon steel	10,800
Subtotal			2,758,500

Area 6--Reheat

Item	No.	Description	Total equipment cost, 1979 $
1. Reheater	2	Indirect steam, shell, carbon steel tubes, 316L, 10,000 ft^2	1,073,000

(continued)

TABLE 30. (continued)

Item	No.	Description	Total equipment cost, 1979 $
2. Soot blowers	20	Air, fixed	130,000
Subtotal			1,203,000

Area 7--Gypsum Production

Item	No.	Description	Total equipment cost, 1979 $
1. Tank, centrifuge feed	2	22 ft dia x 22 ft high, 34,300 gal, open top, flakeglass-lined, carbon steel	99,400
2. Agitator, centrifuge feed tank	2	15 hp, resin-lined, carbon steel	43,100
3. Pump, centrifuge feed	3	Centrifugal belt drive, 510 gpm, 50 ft head, 15 hp, resin-lined, carbon steel	16,700
4. Separator, gypsum	18	Centrifuge, vertical basket, resin-lined, carbon steel	966,600
5. Conveyor, sludge to mixing tank	2	Belt, 14 in. wide x 100 ft long, 2 hp, 35 ft lift, 18.5 tons/hr	51,800
6. Pump, gypsum centrate to tank	3	Centrifugal belt drive, 450 gpm, 50 ft head, 10 hp, resin-lined, carbon steel	15,900
7. Tank, separated liquor	2	27.5 ft dia x 29 ft high, 64,800 gal, open top, flakeglass-lined, carbon steel	162,200
8. Pump, separated liquor to limestone preparation	3	Centrifugal belt drive, 500 gpm, 100 ft head, 25 hp, resin-lined, carbon steel	18,200

(continued)

TABLE 30. (continued)

Item	No.	Description	Total equipment cost, 1979 $
9. Pump, reactor feed	3	Centrifugal belt drive, 50 gpm, 50 ft head, 1.5 hp, resin-lined, carbon steel	7,100
10. Tank, surge	2	45 ft dia x 48 ft high, 571,000 gal, concrete	330,000
11. Pump, surge tank recycle	3	Centrifugal belt drive, 50 gpm, 50 ft head, 1.5 hp, resin-lined, carbon steel	7,100
Subtotal			1,718,100

Area 8--Nitrogen Treatment

Item	No.	Description	Total equipment cost, 1979 $
1. Reactor, gypsum production	2	15 ft dia x 17 ft high, 33,000 ft^3, resin-lined, carbon steel	52,000
2. Agitator, gypsum production reactor	2	10 hp, resin-lined, carbon steel	30,600
3. Pump, reactor effluent to centrifuge feed tank	3	Centrifugal belt drive, 50 gpm, 50 ft head, 1.5 hp, resin-lined, carbon steel	7,000
4. Tank, centrifuge feed	2	10 ft dia x 11 ft high, 3,300 gal, open top, resin-lined, carbon steel	21,300
5. Agitator, centrifuge feed tank	2	7.5 hp, resin-lined, carbon steel	26,000
6. Pump, gypsum centrifuge feed	3	Centrifugal belt drive, 50 gpm, 50 ft head, 1.5 hp, resin-lined, carbon steel	7,000

(continued)

TABLE 30. (continued)

Item	No.	Description	Total equipment cost, 1979 $
7. Separator, gypsum	4	Centrifuge, vertical basket, resin-lined, carbon steel	214,800
8. Conveyor, gypsum to mixing tank	2	Belt, 14 in. wide x 100 ft long, 0.5 hp, 35 ft lift, 3 tons/hr	51,200
9. Pump, centrate from centrifuge to evaporator feed tank	3	Centrifugal belt drive, 50 gpm, 50 ft head, 1.5 hp, resin-lined, carbon steel	7,000
10. Pump, waste wash H_2O to evaporator feed tank	3	Centrifugal belt drive, 100 gpm, 50 ft head, 3.0 hp, resin-lined, carbon steel	9,800
11. Tank, evaporator feed	2	10 ft dia x 11 ft high, 3,300 gal, open top, resin-lined, carbon steel	21,300
12. Pump to evaporators	3	Centrifugal belt drive, 50 gpm, 50 ft head, 1.5 hp, resin-lined, carbon steel	7,000
13. Evaporator	2	Double effect, vacuum, 8.9 ft dia x 23 ft high, resin-lined, carbon steel	991,000
14. Pump, fluid fertilizer storage feed	3	Centrifugal belt drive, 15 gpm, 50 ft head, 0.5 hp, resin-lined, carbon steel	6,800
15. Tank, fluid fertilizer storage	1	7.5 ft dia x 10 ft high, 3,300 gal, closed top, resin-lined, carbon steel	8,200
16. Pump, feed to bulk fertilizer storage	2	Centrifugal belt drive, 30 gpm, 150 ft head, 3 hp, resin-lined, carbon steel	9,600

(continued)

TABLE 30. (continued)

Item	No.	Description	Total equipment cost, 1979 $
17. Tank, bulk fluid fertilizer storage	1	35 ft dia x 40 ft high, 287,900 gal, closed top, resin-lined, carbon steel	164,200
Subtotal			1,625,800

Area 9--Raw Material Storage

Item	No.	Description	Total equipment cost, 1979 $
1. Elevator, catalyst to primary silo	1	Bucket, 6 in. x 4 in. x 4-1/4 in., 0.25 hp, 35 ft lift, 7 tons/hr	6,900
2. Catalyst silo	1	24 ft dia x 24 ft high, closed top, flakeglass-lined, carbon steel	71,000
3. Weigh feeder	1	0.5 tons/hr	6,300
4. Conveyor catalyst to tank	1	Belt, 14 in. wide x 100 ft long, 0.25 hp, 50 ft lift, 0.5 tons/hr	27,100
5. Catalyst silo	1	5 ft dia x 5 ft high, closed top, PVC	1,200
6. Weigh feeder	1	16 lb/hr	6,300
7. Elevator, $(NH_4)_2SO_4$ to silo	1	Bucket, 6 in. x 4 in. x 4-1/2 in., 2 hp, 50 ft lift, 13 tons/hr	8,900
8. Silo, $(NH_4)_2SO_4$	1	40 ft dia x 42.5 ft high, carbon steel	203,100
9. Conveyor, $(NH_4)_2SO_4$ to intermediate surge	2	Belt, 14 in. wide x 300 ft long, 1.5 hp, 30 ft lift, 4.1 tons/hr	148,400

(continued)

TABLE 30. (continued)

Item	No.	Description	Total equipment cost, 1979 $
10. Silo $(NH_4)_2SO_4$	2	10 ft dia x 13 ft high, carbon steel	28,300
11. Weigh feeder	2	2 tons/hr	12,600
12. Conveyor, $(NH_4)_2SO_4$ to reactor	2	Belt, 14 in. wide x 100 ft long, 1 hp, 30 ft lift, 2 tons/hr	50,000
Subtotal			570,100

Area 10--Sludge Disposal

Item	No.	Description	Total equipment cost, 1979 $
1. Tank, mixing	1	27.5 ft dia x 30 ft high, 133,300 gal, open top, resin-lined, carbon steel	83,500
2. Agitator, mixing tank	1	40 hp, resin-lined, carbon steel	43,900
3. Pump, sludge to pond	2	Centrifugal belt drive, 520 gpm, 300 ft head, 75 hp, resin-lined, carbon steel	35,400
4. Sludge pond	1	253 acres	5,882,800
5. Pump, return from pond	2	Centrifugal belt drive, 350 gpm, 200 ft head, 30 hp, resin-lined, carbon steel	25,200
Subtotal			6,070,800

(continued)

TABLE 30. (continued)

Area 11--Chlorine Dioxide Generation

	Item	No.	Description	Total equipment cost, 1979 $
1.	Chlorine dioxide generation system	2	22 tons/day capacity	15,709,000
	Subtotal			15,709,000
	Total, Areas 1-11			33,871,700

Particulate Control--

A particulate control efficiency of 99.5% is achieved by using a 94% efficient ESP followed by a prescrubber. (The ESP description and cost are included later in the ESP section.) Although only a single ESP is included, the flue gas duct is then split and two trains of fans, prescrubbers, and absorbers are included. The prescrubber is a Moretana plate tower. The purge stream is neutralized in pH-controlling tanks and pumped to the fly ash pond.

SO_x-NO_x Absorption--

Simultaneous SO_x-NO_x absorption is carried out in two trains of Moretana plate towers. Absorbent recycle tanks and pumps are also provided. In addition, the Moretana Calcium process has a separate loop between the absorber and the reheater which includes a clarifier and mist eliminator to control liquid carryover.

Reheat--

Flue gas reheat to 175°F is provided by two reheaters using an indirect steam system.

Gypsum Production--

In the Moretana Calcium process no attempt is made to further treat the mixed sulfite-sulfate salts and the mixture is simply separated, reslurried, and pumped to the sludge pond.

Nitrogen Treatment Section--

The use of ClO_2 as a gas-phase oxidant results in the conversion of 50% of the absorbed NO_x to nitrate salts. Since waste water nitrate salts could result in a significant water pollution problem, a purge stream is further treated with $(NH_4)_2SO_4$ to form additional gypsum sludge and ammonium nitrate (NH_4NO_3), a potential fertilizer material. The NH_4NO_3 solution is concentrated in one to two double-effect evaporators and stored as a byproduct.

Raw Material Storage--

This section includes all the processing equipment needed to receive, store, and retrieve for later use all of the various raw materials (except limestone) needed in the process. Storage capacity is designed for 30 days of normal usage.

Sludge Disposal--

All the equipment needed to reslurry the mixed sulfite-sulfate sludge from the gypsum production section and the gypsum from the nitrogen treatment section is included. In addition, a 253-acre sludge pond is included.

ClO_2 Generation--

Two separate ClO_2 generating systems are provided. Each has a 22-ton/day capacity and is designed for treating the flue gas for one absorber. Provisions have been made for temporary in-process storage of ClO_2 such that fluctuations in the NO_x concentration in the flue gas can be handled effectively.

HITACHI ZOSEN NO_x-ONLY PROCESS - DRY, SCR

Process Description

Hitachi Zosen (Hitachi Shipbuilding and Engineering Company, Ltd.) has developed an NO_x FGT process for dry, SCR of NO_x with NH_3 (8). Hitachi Zosen has developed a catalyst and reactor design which permits treatment of flue gas with a high-particulate loading. Therefore, the flue gas from a coal-fired boiler may be fed directly from the economizer to the reactor, upstream from the air heater, without any particulate removal treatment. Before entering the reactor the flue gas is injected with NH_3 which has been diluted with air to a 5% NH_3 concentration to enhance mixing. In the reactor NO_x is reduced to nitrogen by reaction with NH_3 in the presence of a catalyst at a temperature range of 572-752°F. Hitachi Zosen reports that the level of excess NH_3 in the flue gas leaving the system is very low (<20 ppm). The treated flue gas is passed through the boiler air heater.

The NH_3 to NO_x mol ratio employed for $\geq$90% removal efficiency is about (1.0-1.2):1. Hitachi Zosen claims the pressure drop across the reactor is 2-3 inches water. The area velocity (flow rate of gas/surface area of catalyst) is between 0.38-0.55 $ft^3/min/ft^2$ which is reported to correspond to a space velocity (flow rate of gas/volume of catalyst) of 5,000-10,000 $ft^3/hr/ft^3$. The process is reported to be capable of functioning with particulate loading up to about 7 gr/sft^3, making it possible to use this system on coal-fired flue gas before particulate removal.

The NH_3 flow rate is automatically controlled based on the flue gas rate to the reactor, reactor inlet and outlet NO_x concentration, and NH_3 outlet concentration. To prevent formation of $(NH_4)_2SO_4$ the catalyst bed temperature can be controlled by bypassing a part of the high-temperature flue gas flow around the economizer to the reactor.

The catalyst is manufactured in the shape of corrugated units. These units are joined together for the particular size required. The flue gas passes parallel to the catalyst surface. The catalyst composition has not been revealed for proprietary reasons; however, Hitachi Zosen does state that it is constructed of common material. The catalyst life is guaranteed for 1 year but the actual life may be higher.

Analysis of Processing Subsections

The flow diagram and material balance for the base case are shown in Figure 5 and Table 31 respectively. The Hitachi Zosen process is divided into three processing sections and the equipment assigned to the appropriate section. The equipment list and descriptions by area are presented in Table 32. The total land requirement is estimated to be 1.5 acres.

NH_3 Storage and Injection--

A compressor (and spare) for unloading liquid NH_3 from truck or rail transport and 250 psig storage tanks for a 30-day supply are included. Before NH_3 is injected into the flue gas, it is vaporized in a shell-and-

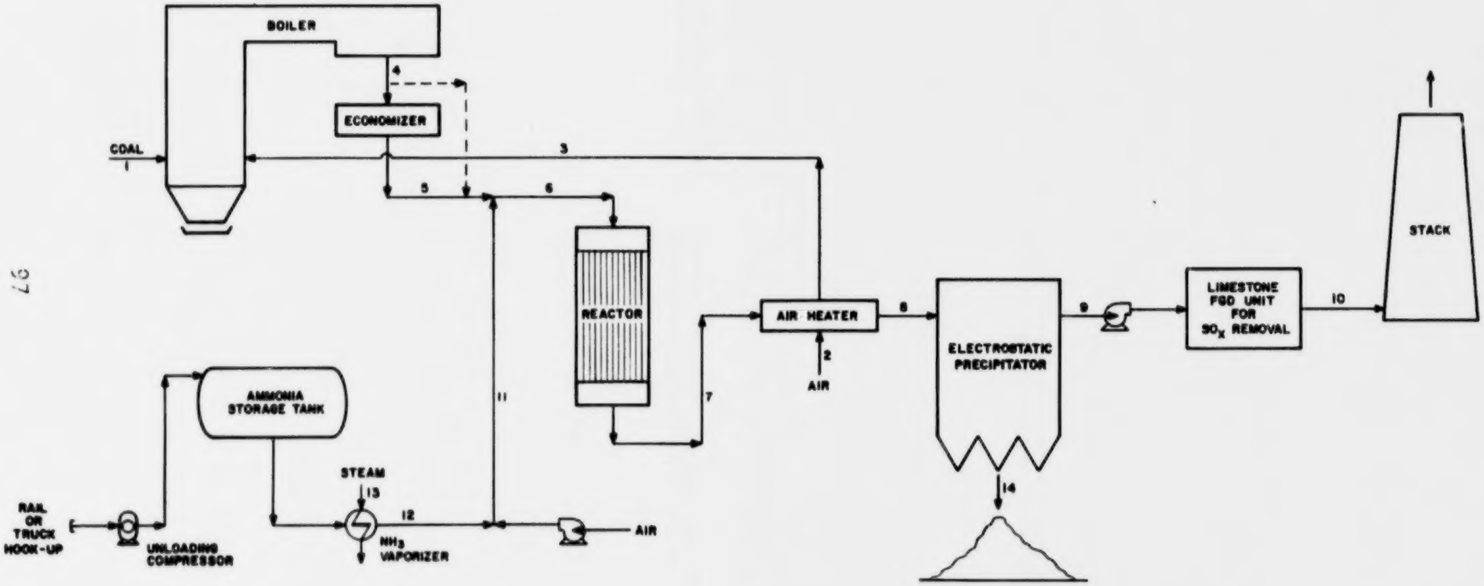
BOILER
COAL
1
4
ECONOMIZER
3
5
6
REACTOR
7
AIR HEATER
2
AIR
8
ELECTROSTATIC PRECIPITATOR
14
9
LIMESTONE FGD UNIT FOR SO_x REMOVAL
10
STACK
11
AMMONIA STORAGE TANK
RAIL OR TRUCK HOOK-UP
UNLOADING COMPRESSOR
STEAM
13
12
NH_3 VAPORIZER
AIR

Figure 5. Hitachi Zosen process. Flowsheet.

TABLE 31. HITACHI ZOSEN NO_x-ONLY PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Flue gas to reactor
1	Total stream, lb/hr	428,600	4,546,200	4,101,800	4,516,100	4,516,100
2						
3	Sft^3/min (60°F)		1,005,000	906,700	958,000	958,000
4	Temperature, °F		80	535	890	705
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	6	7	8	9	10
	Description	Flue gas-NH_3 mixture to reactor	Gas to air heater	Gas to ESP	Gas to FGD unit	Treated flue gas to stack
1	Total stream, lb/hr	4,576,700	4,576,700	5,021,000	4,966,500	5,165,800
2						
3	Sft^3/min (60°F)	971,800	971,800	1,070,200	1,070,200	1,143,400
4	Temperature, °F	697	712	300	300	175
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	11	12	13	14	
	Description	NH_3 stream injected into flue gas	NH_3 from storage	Steam to vaporizer	Fly ash from ESP	
1	Total stream, lb/hr	60,600	1,827	1,120	54,600	
2						
3	Sft^3/min (60°F)	13,800				
4	Temperature, °F		110	298		
5	Pressure, psig		225	50		
6						
7						
8						
9						
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

TABLE 32. HITACHI ZOSEN NO_x-ONLY PROCESS

BASE CASE EQUIPMENT LIST
DESCRIPTION AND COST

Area 1--NH_3 Storage and Injection

Item	No.	Description	Total equipment cost, 1979 $
1. Compressor, NH_3 unloading	2	Single cylinder, double acting, 300 sft^3/min at 250 psig, 30 psig suction, 125 hp, cast iron	50,000
2. Tank, NH_3 storage	8	Horizontal, 9 ft dia x 66 ft long, 30,000 gal, 250 psig, carbon steel	230,100
3. Vaporizer, NH_3	2	Steam at 298°F, tube type, 30 ft^2, 1.03 MBtu/hr, carbon steel	36,800
4. Blower, NH_3 and air	2	7,000 aft^3/min, ΔP 5 in. H_2O, 10 hp, carbon steel	13,500
Subtotal			330,400

Area 2--Reactor Section

Item	No.	Description	Total equipment cost, 1979 $
1. Reactor	2	40 ft x 40 ft x 31 ft high, 715°F operating temperature, carbon steel, insulated, with fly ash hoppers and sootblowers	1,029,000
2. Initial catalyst fill		Proprietary catalyst, manufactured in honeycomb-shaped units	6,370,000
Subtotal			7,399,000

(continued)

TABLE 32. (continued)

Area 3--Flue Gas Fans				
	Item	No.	Description	Total equipment cost, 1979 $
1.	Blower, flue gas	4	391,025 aft^3/min, ΔP 22 in. H_2O, 2,250 hp, 316 stainless steel	405,000
	Subtotal			405,000
	Total, Areas 1-3			8,134,400

tube steam-heated vaporizer and mixed with air supplied by a small blower to form a 5% NH_3-in-air mixture, since the inflammability limits for NH_3 in air are 15.5 to 27.0%, and for easier flow control and for mixing enhancement. There is one vaporizer and one air-NH_3 blower for each of the two reactor trains. The assumed pressure drop for sizing the NH_3-air blower is 5 inches water.

The NH_3 injection grid is primarily piping and the cost therefore is incorporated into the area investment estimate shown in the summary capital investment estimates in the following report section, Economic Evaluation and Comparison.

Reactor Section--

Two reactors are provided, each handling 50% of the total flow. The reactors are fixed-bed type and constructed of carbon steel. Each reactor is provided with fly ash hoppers for collection of any deposited fly ash and with two soot blowers for periodic cleaning. The catalyst is manufactured into corrugated shaped units and these units are joined together to form the appropriately sized catalyst bed.

Gas Handling--

A larger ID blower is provided downstream of each ESP to compensate for the pressure drop created by the boiler, FGT system, and ESP. However, only the incremental cost attributed to the NO_x removal system is included in the cost estimates. Because of the larger pressure drop, additional costs to prevent implosion are included with gas handling.

KURABO KNORCA NO_x-ONLY PROCESS - DRY, SCR

Process Description

The Kurabo Knorca process is based on SCR of NO_x with NH_3 using a moving-bed catalyst developed by Kurabo (5). Two types of catalyst systems have been developed by Kurabo. One system uses copper oxide on an aluminum oxide carrier which requires thermal regeneration after each pass through the reactor. The other system is an iron-oxide catalyst on a titanium dioxide carrier which, according to the process developers, does not require thermal regeneration and is more applicable to treatment of coal-fired flue gas and for gases in which chlorides exceed 100 ppm. The latter catalyst is used for this evaluation.

The Knorca reactor removes about 80% of the fly ash in the flue gas but it cannot be operated at the full-fly ash loading of coal-fired flue gas. Therefore, a 98% efficient hot ESP is used between the economizer and reactor to reduce the fly ash level in the flue gas going to the reactor to 0.44 gr/sft^3. The overall particulate removal efficiency of the system is 99.5%. The flue gas leaving the ESP is injected with NH_3 and passed through a static gas mixer to disperse the NH_3 in the flue gas before it enters the reactor. The reactor contains a bed in which the spherical catalyst (3-5 mm diameter) moves slowly and continuously downward as the flue gas passes through it in crossflow. The NO_x is reduced to molecular nitrogen and water in the presence of the catalyst. The flue gas leaving the reactor continues through the boiler air heater. Fly ash and broken catalyst are separated from the catalyst removed from the reactor by a vibrating sieve, and the catalyst is conveyed by horizontal vibrating conveyors and bucket elevators to the reactor feed hopper. Makeup catalyst is manually added to replace the broken catalyst. A baghouse and air blower are used to collect the fly ash separated from the catalyst in the vibrating sieve.

The reactor design operating conditions are 716°F, atmospheric pressure, and a space velocity of 7,500 $ft^3/hr/ft^2$. The flue gas residence time is about 0.15 second in the catalyst bed and 2-3 seconds in the reactor. The pressure drop across the reactor and static gas mixer is 4 inches water and 2 inches water respectively. The catalyst panels in each reactor are divided into modules of 10- to 15-MW individual capacity. The expected catalyst life is at least 2 years and the loss makeup rate is about 10%/year.

The NH_3 flow rate is automatically controlled on the basis of the inlet gas NO_x concentration and flue gas rate. As an additional control, the NH_3 usage rate is also compared with fuel consumption. Flue gas temperature can be controlled by passing a part of the flue gas stream around the economizer.

TRW, Inc., has exclusive rights to supply the Knorca system for boilers in the United States and Canada.

Analysis of Processing Subsections

The flow diagram and material balance are shown in Figure 6 and Table 33 respectively. The Knorca process is divided into four processing sections and the equipment assigned to the appropriate section. Table 34 shows the equipment list and description by area. The total land requirement is estimated to be 1.5 acres.

NH_3 Storage and Injection--

A compressor (including a spare) for unloading liquid NH_3 from truck or rail transport and 250 psig storage tanks sufficient to maintain a 30-day supply are included. Four steam-heated, shell-and-tube vaporizers are provided, one for each reactor train. The NH_3 injection grid is primarily piping and the cost is therefore incorporated into the area investment estimates shown in the following report section, Economic Evaluation and Comparison.

Reactor Section--

A static gas mixer is used to ensure thorough dispersion of the NH_3 in the flue gas prior to the mixture entering the reactor. There are four reactor trains, as recommended by Kurabo, and each is equipped with a catalyst processing system. Each reactor contains 92.6 tons of the iron-based catalyst. The catalyst panels are divided into modules of 10- to 15-MW equivalent capacity. The reactor housing is carbon steel, but the screen which interfaces the flue gas and catalyst is stainless steel. The sieve which separates fly ash and broken catalyst from the catalyst is carbon steel construction. The recycled catalyst is transported to a bucket elevator by a horizontal vibratory conveyor sized for 0.53 ton/hr. The catalyst is

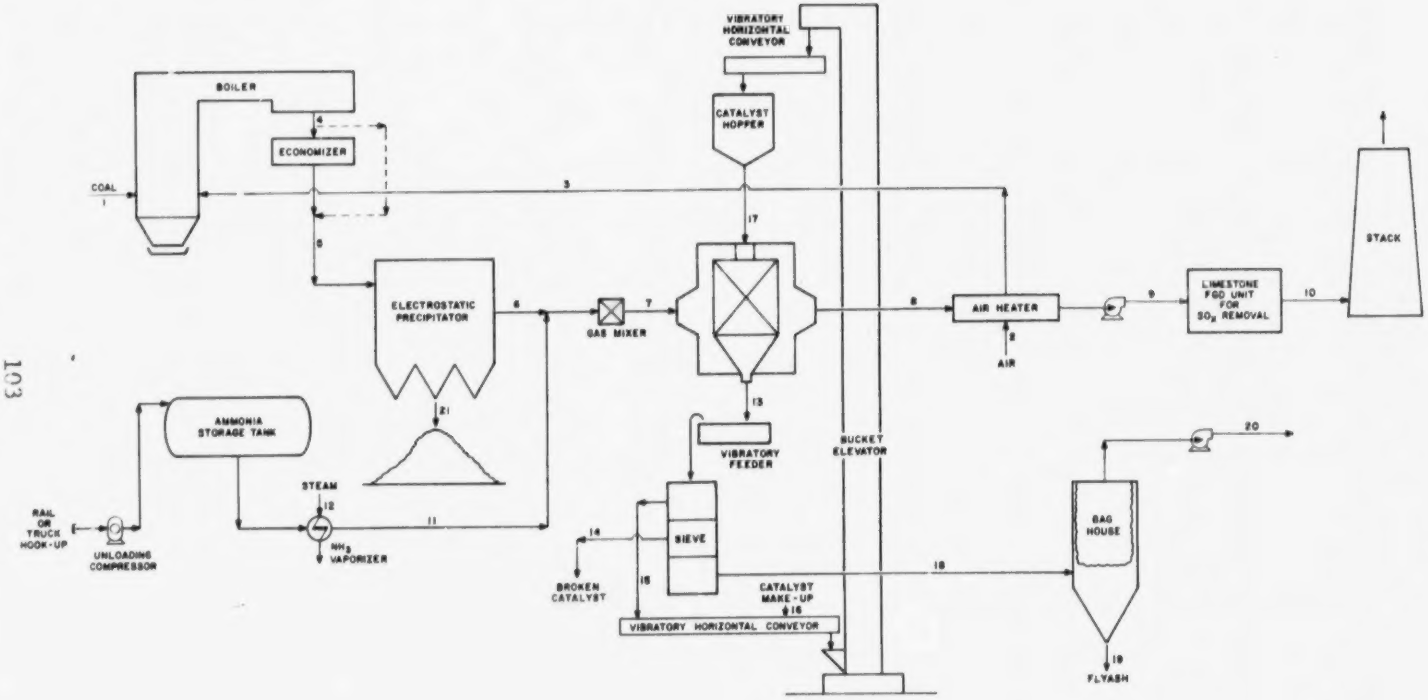
COAL
1
BOILER
4
ECONOMIZER
3
5
ELECTROSTATIC PRECIPITATOR
21
6
GAS MIXER
7
VIBRATORY HORIZONTAL CONVEYOR
CATALYST HOPPER
17
13
VIBRATORY FEEDER
BUCKET ELEVATOR
8
AIR HEATER
2
AIR
9
LIMESTONE FGD UNIT FOR SO_x REMOVAL
10
STACK
AMMONIA STORAGE TANK
RAIL OR TRUCK HOOK-UP
UNLOADING COMPRESSOR
STEAM
12
NH_3 VAPORIZER
11
14
BROKEN CATALYST
15
SIEVE
CATALYST MAKE-UP
16
VIBRATORY HORIZONTAL CONVEYOR
18
BAG HOUSE
19
FLYASH
20

Figure 6. Kurabo Knorca process. Flowsheet.

TABLE 33. KURABO KNORCA NO_x-ONLY PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to ESP
1	Total stream, lb/hr	428,600	4,546,200	4,101,800	4,516,100	4,516,100
2						
3	Sft^3/min (60°F)		1,005,000	906,700	958,000	958,000
4	Temperature, °F		80	535	890	705
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	6	7	8	9	10
	Description	Flue gas to gas mixer	Flue gas-NH_3 mixture to reactor	Gas to air heater	Gas to FGD unit	Treated flue gas to stack
1	Total stream, lb/hr	4,462,300	4,463,800	4,462,900	4,907,300	5,106,700
2						
3	Sft^3/min (60°F)	958,000	958,600	958,700	1,057,000	1,130,000
4	Temperature, °F	705	705	717	300	175
5	Pressure, psig					
6						
7						
8						
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10						

	Stream No.	11	12	13	14	15
	Description	NH_3 from storage	Steam for NH_3 vaporization	Catalyst and fly ash to sieve	Broken catalyst	Circulating catalyst
1	Total stream, lb/hr	1,533	940	5,110	10.58	4,222
2						
3	Sft^3/min (60°F)	570				
4	Temperature, °F	110	298			
5	Pressure, psig	225	50			
6						
7						
8						
9						
10						

	Stream No.	16	17	18	19	20
	Description	Catalyst makeup	Reactor catalyst feed	Gas stre[illegible] baghouse	Fly ash exiting baghouse	Air from baghouse
1	Total stream, lb/hr	10.58	4,2[illegible]	29,051	878	28,200
2						
3	Sft^3/min (60°F)			6,280		5,850
4	Temperature, °F					100
5	Pressure, psig					Atm
6						
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10						

(continued)

TABLE 33. (continued)

	Stream No.	21				
	Description	Fly ash from ESP				
1	Total stream, lb/hr	53,800				
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6						
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TABLE 34. KURABO KNORCA NO_x-ONLY PROCESS

BASE CASE EQUIPMENT LIST
DESCRIPTION AND COST

Area 1--NH_3 Storage and Injection

Item	No.	Description	Total equipment cost, 1979 $
1. Compressor, NH_3 unloading	2	Single cylinder, double acting, 300 sft^3/min at 250 psig, 30 psig suction, 125 hp, cast iron	50,000
2. Tank, NH_3 storage	8	Horizontal, 9 ft dia x 66 ft long, 30,000 gal, 250 psig, carbon steel	230,100
3. Vaporizer, NH_3	4	Steam at 298°F, tube type, 15 ft^2, 1.03 MBtu/hr, carbon steel	52,200
Subtotal			332,300

Area 2--Reactor Section

Item	No.	Description	Total equipment cost, 1979 $
1. Static gas mixer	4	13.5 ft x 13.5 ft, minimum length 25 ft, 35,800 aft^3/sec at 705°F, carbon steel	86,400
2. Reactor	4	21.3 ft wide x 82 ft long x 42.7 ft high, 715°F operating temperature, atmospheric pressure, space velocity = 7,500 hr^{-1}, housing is carbon steel, screen is stainless steel	1,996,800
3. Feeder, reactor to sieve	4	Vibratory, 0.64 tons/hr, 21.8 ft^3/hr, 0.25 hp	5,600

(continued)

TABLE 34. (continued)

Item	No.	Description	Total equipment cost, 1979 $
4. Sieve	4	0.64 tons/hr, 21.8 ft^3/hr, 1 hp, screen and connector are stainless steel	20,000
5. Conveyor, sieve to bucket elevator	4	Vibratory horizontal, 82 ft long (max), 0.53 tons/hr, 1 hp, trough = 9 in. x 2 in.	19,400
6. Elevator, circulating catalyst to catalyst reservoir	4	Continuous, bucket = 6 in. x 4 in., 130 ft high (max), 0.53 tons/hr, 17 ft^3/hr, 2 hp	48,400
7. Conveyor, bucket elevator to catalyst hopper	4	Vibratory horizontal, 82 ft long (max), 0.53 tons/hr, 1 hp, trough = 8 in. x 2 in.	19,400
8. Hopper, catalyst	4	7.5 ft dia x 8 ft high, 353 ft^3 (without conical bottom) conical bottom = 6.5 ft high, 449 ft^3 total, carbon steel	23,500
9. Initial catalyst fill	4	92.6 tons/train, Fe-based catalyst on TiO_2 support	3,936,300
Subtotal			6,155,800

Area 3--Waste Disposal

Item	No.	Description	Total equipment cost, 1979 $
1. Baghouse filter	4	1,666 aft^3/min, 220 lb/hr, installed cloth area = 833 ft^2, operating cloth area = 666 ft^2	32,000

(continued)

TABLE 34. (continued)

Item	No.	Description	Total equipment cost, 1979 $
2. Blower, baghouse	4	1,666 aft^3/min, ΔP 4 in. H_2O, 2 hp, carbon steel	13,400
Subtotal			45,400

Area 4--Flue Gas Fans

Item	No.	Description	Total equipment cost, 1979 $
1. Blower, flue gas	4	386,000 aft^3/min, ΔP 22 in. H_2O, 2,250 hp, 316 stainless steel	405,000
Subtotal			405,000
Total, Areas 1-4			6,938,500

transporated by another vibratory conveyor from the bucket elevator to the catalyst hopper which feeds the reactor. The catalyst hopper is carbon steel and has a 449 ft^3 volume.

Waste Disposal--

The fly ash removed from the catalyst at the sieve is collected by a baghouse filter for disposal. The baghouse filter and baghouse blower are sized on the basis of: (1) fly ash loading of 40 gm/Nm^3, (2) operating air/cloth ratio of 2.5 $aft^3/min/ft^2$ cloth, (3) installed air/cloth ratio of 2.0 $aft^3/min/ft^2$ cloth, and (4) pressure drop of 4 inches water. There is one baghouse filter and blower per reactor train.

Gas Handling--

Four higher energy ID blowers are provided downstream from the air heater to compensate for the pressure drop created by the NO_x removal system. Only the incremental costs attributed to the increased pressure drop of the NO_x removal system are included in the estimates. A pressure drop of 7 inches water for the NO_x removal system is used as a basis. Because of the larger pressure drop, additional costs to prevent implosion are included with gas handling.

UOP SFGT-N, NO_x-ONLY PROCESS - DRY, SCR

Process Description

In addition to simultaneous SO_x-NO_x removal the UOP process can be used as an NO_x-only process (21). The same catalyst--copper sulfate on an aluminum oxide carrier--as used in the SO_x-NO_x process is used but the catalyst is not regenerated. When the copper oxide is converted to copper sulfate by SO_x in the flue gas, it remains in this form and the SO_x passes unchanged through the system. Flue gas leaving the economizer is injected with an NH_3-air mixture and passed through a fixed-bed, parallel-passage reactor where NO_x is reduced to molecular nitrogen and water by the NH_3 in the presence of the catalyst. The NH_3 is diluted with air to a 5% by volume NH_3-air mixture before injection to ensure adequate mixing with the flue gas.

The reactor operating conditions are similar to those of the UOP SFGT-SN SO_x-NO_x removal process:

Maximum particulate loading	Full loading (>10 gr/sft^3)
Pressure drop across the reactor	5-6 inches water
NH_3:NO_x mol ratio	(1.0-1.2):1
Average space velocity	4,000-8,000 $ft^3/hr/ft^2$
Temperature	750°F

With the NO_x FGT system located between the economizer and air heater, the recommended method of controlling flue gas reaction temperature consists of regulating a bypass flow of either flue gas or boiler feedwater around the economizer to control the heat transfer rate. Also, all the reactions occurring in the reactor are exothermic and by having the FGT system located upstream of the air heater this heat of reaction can be recovered from the flue gas.

The parallel-passage reactor has been shown to be capable of operating at normal conditions with full particulate loading (>10 gr/sft^3), although reactor cleaning methods were necessary in some cases. Initial testing with full-fly ash loading to a pilot-plant reactor at a test unit in Pernis, The Netherlands, showed no deterioration in performance. However, at a 40-MW equivalent facility at Showa Yokkaichi Sekiyo refinery in Yokkaichi, Japan, the high vanadium and sodium content of the fly ash gradually fouled the reactor internals, limiting operating periods to 1-2 months. Modifications recently have allowed an operating period of approximately 1 year without significant deterioration in performance. The different particulate loading with coal-fired flue gas (>10 gr/sft^3) at a test facility on a Tampa Electric Company boiler caused a decline in performance after only days of operation. As a result of this problem, a procedure was implemented at this pilot plant which provided in situ cleaning of reactor internals during normal operation. This technique allowed stable performance at high-particulate loadings without any adverse effects on the reactor pressure drop or on the FGD system.

Shell International Petroleum Maatschappij developed this process and has been involved in the program since the early 1960's. The UOP Process Division holds the worldwide licensing rights to the SFGT process (with the exception of the Far East).

Analysis of Processing Subsections

The flow diagram and material balance are shown in Figure 7 and Table 35 respectively. The UOP SFGT-N process is divided into three sections and the process equipment assigned to the appropriate section. Table 36 shows the equipment list and description by area. The total land requirement is estimated to be 1.5 acres.

NH_3 Storage and Injection--

A compressor, including a spare, for unloading liquid NH_3 from truck or rail transport and storage tanks for a 30-day supply of NH_3 at 250 psig are included. A steam-heated, shell-and-tube vaporizer is provided for vaporization of NH_3, prior to its being diluted with air. The air for dilution of the NH_3 is supplied by a blower sized for a 5-inch water pressure drop. There is a vaporizer and blower for each of the two reactor trains. The NH_3 injection grid is primarily piping and the cost is incorporated into the area investment estimate shown in the summary capital investment estimates in the following report section, Economic Evaluation and Comparison.

Reactor Section--

Two reactors are provided, each of which handles 50% of the flue gas and NH_3 mixture. The reactors are of parallel-passage design and made of carbon steel. The catalyst is contained within unit cells and the flue gas is forced across the face of the catalyst layers, not through the catalyst.

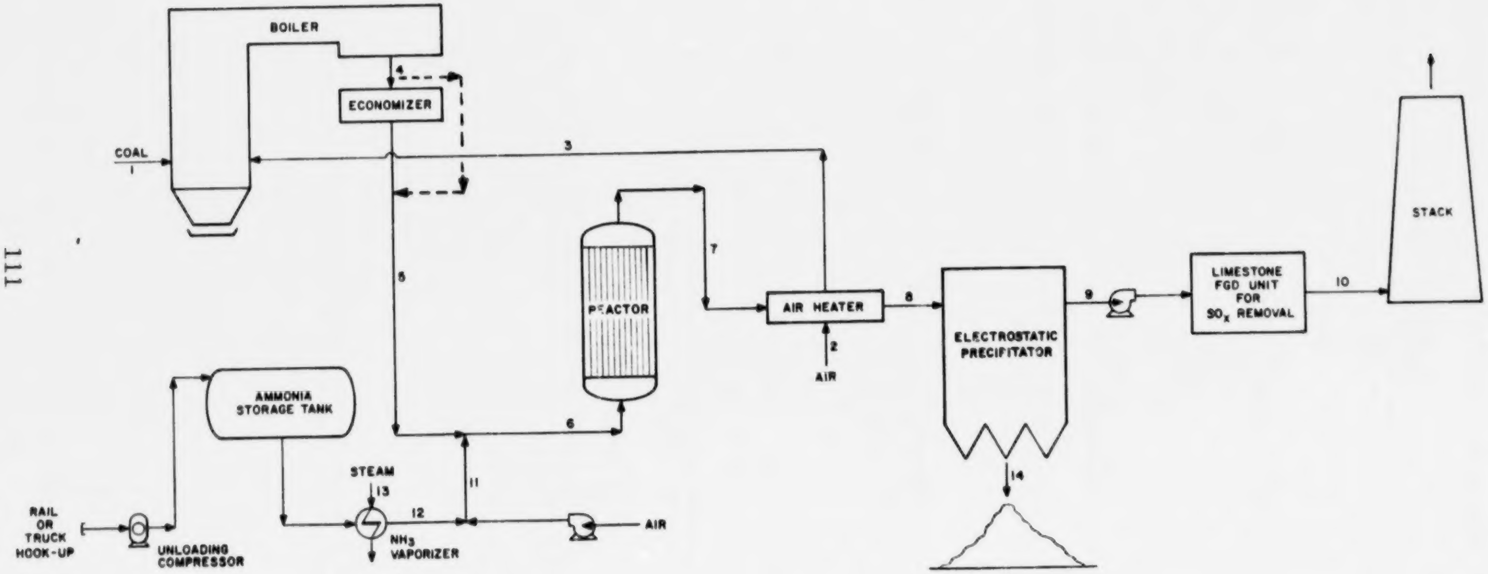
BOILER
4
ECONOMIZER
COAL
1
3
5
REACTOR
7
AIR HEATER
2
AIR
8
ELECTROSTATIC PRECIPITATOR
14
9
LIMESTONE FGD UNIT FOR SO_x REMOVAL
10
STACK
AMMONIA STORAGE TANK
6
STEAM
13
11
12
AIR
RAIL OR TRUCK HOOK-UP
UNLOADING COMPRESSOR
NH_3 VAPORIZER

Figure 7. UOP SFGT-N, NO_x-only process. Flowsheet.

TABLE 35. UOP SFGT-N, NO_x-ONLY PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Flue gas to reactor
1	Total stream, lb/hr	428,600	4,546,200	4,101,800	4,516,100	4,516,100
2						
3	Sft^3/min (60°F)		1,005,000	906,700	958,000	958,000
4	Temperature, °F		80	535	890	705
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	6	7	8	9	10
	Description	Flue gas-NH_3 mixture to reactor	Gas to air heater	Gas to ESP	Gas to FGD unit	Treated flue gas to stack
1	Total stream, lb/hr	4,576,700	4,576,700	5,021,000	4,966,400	5,168,700
2						
3	Sft^3/min (60°F)	971,800	971,800	1,070,200	1,070,200	1,143,700
4	Temperature, °F	697	212	300	300	175
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	11	12	13	14	
	Description	NH_3 stream injected into flue gas	NH_3 from storage	Steam to vaporizer	Fly ash from ESP	
1	Total stream, lb/hr	60,600	1,827	1,120	54,600	
2						
3	Sft^3/min (60°F)	13,800				
4	Temperature, °F		110	298		
5	Pressure, psig		225	50		
6						
7						
8						
9						
10						

TABLE 36. UOP SFGT-N, NO_x-ONLY PROCESS

BASE CASE EQUIPMENT LIST
DESCRIPTION AND COST

Area 1--NH_3 Storage and Injection

	Item	No.	Description	Total equipment cost, 1979 $
1.	Compressor, NH_3 unloading	2	Single cylinder, double acting, 300 sft^3/min at 250 psig, 30 psig suction, 125 hp, cast iron	50,000
2.	Tank, NH_3 storage	8	Horizontal, 9 ft dia x 66 ft long, 30,000 gal, 250 psig, carbon steel	230,100
3.	Vaporizer, NH_3	2	Steam at 298°F, tube type, 30 ft^2, 1.03 MBtu/hr, carbon steel	36,800
4.	Blower, NH_3 and air	2	7,000 aft^3/min, ΔP, 5 in. H_2O, 10 hp, carbon steel	13,500
	Subtotal			330,400

Area 2--Reactor Section

	Item	No.	Description	Total equipment cost, 1979 $
1.	Reactor	2	39 ft dia x 24 ft long, 715°F operating temperature, pressure same as flue gas duct, carbon steel, insulated Included is a prorated allowance for a UOP-SFGT package consisting of reactor internals with initial catalyst fill, process engineering, license fees, and startup assistance	5,428,000
	Subtotal			5,428,000

(continued)

TABLE 36. (continued)

Area 3--Flue Gas Fans

Item	No.	Description	Total equipment cost, 1979 $
1. Blower, flue gas	4	391,027 aft^3/min, ΔP 22 in. H_2O, 2,250 hp, 316 stainless steel	405,000
Subtotal			405,000
Total, Areas 1-3			6,163,400

Gas Handling--

A larger ID blower is provided downstream of each ESP to compensate for the pressure drop created by the boiler, FGT system, and ESP. However, only the incremental cost attributed to the NO_x removal system is included in the cost estimates. A pressure drop of 7 inches water for the NO_x process is used as a basis for the incremental cost calculations. Because of the larger pressure drop, additional costs to prevent implosion are included with gas handling.

UOP SFGT-SN, SO_x-NO_x PROCESS - DRY, SCR

Process Description

UOP offers a dry, simultaneous SO_x-NO_x removal system using the Shell CuO process (21). The SO_2 in the flue gas reacts with the sorbent Cuo and forms copper sulfate ($CuSO_4$) which acts as a catalyst in the reduction of NO_x with NH_3. Upon catalyst regeneration a concentrated SO_2 off-gas stream is produced. This SO_2 off-gas is then further processed to manufacture byproduct H_2SO_4.

The flue gas leaving the boiler economizer is injected with NH_3 and is passed through a reactor in the acceptance stage. The catalyst is contained in a fixed-bed, parallel-passage reactor with the flue gas flowing across the face of the catalyst, not through it. The catalyst is copper on a special aluminum oxide support. The catalyst is rapidly oxidized upon contact with the flue gas by the following reaction:

$$Cu_{(s)} + 1/2O_{2(g)} \rightarrow CuO_{(s)} \qquad (1)$$

The SO_2 in the flue gas reacts with this sorbent as shown below:

$$SO_{2(g)} + 1/2O_{2(g)} + CuO_{(s)} \rightarrow CuSO_{4(s)} \qquad (2)$$

In the presence of this $CuSO_4$ the NO_x in the flue gas is reduced to molecular nitrogen and water by NH_3. The treated flue gas then leaves the reactor and passes through the steam superheater, the boiler air heater, and a 99.5% efficient "cold" ESP before going to the stack.

When the acceptor material becomes saturated such that the SO_2 removal efficiency drops below 90% removal, the flue gas flow is diverted to a reactor containing regenerated acceptor material. The spent acceptor is regenerated with a steam-diluted, hydrogen-containing regeneration gas which is produced by steam-reforming naphtha. (There are other sources of hydrogen which could be considered, e.g., coal gasification.) Since this regeneration occurs at the same temperature as acceptance, heating or cooling of the acceptor is not required. The $CuSO_4$ is converted to elemental copper (Cu), SO_2, and water, while any unreacted CuO is reduced to Cu and water. The following equations express the above reactions:

$$CuSO_{4(s)} + 2H_2O_{(g)} \rightarrow Cu_{(s)} + SO_{2(g)} + 2H_2O_{(g)} \qquad (3)$$

$$CuO_{(s)} + H_{2(g)} \rightarrow Cu_{(s)} + H_2O_{(g)} \qquad (4)$$

The off-gas produced during regeneration contains SO_2, water vapor, and traces of unreacted reducing gas and may be handled by various means depending on the desired product and the inerts in the reducing gas. For this study this off-gas is sent to a captive sulfuric acid plant to produce acid. (An alternative SO_2 processing scheme which could be considered is the generation of elemental sulfur from a Claus unit.)

Since this FGT process is cyclic, the off-gas production is also cyclic and before further processing of the regeneration off-gas flow variations must be minimized. The bulk of the water vapor also must be condensed. Therefore, the off-gas passes through a waste heat boiler to recover sensible heat and a direct-contact cooler for cooling below the dewpoint. A gas compressor and gas holder are considered the simplest system for dampening the flow.

Normal operating conditions for the UOP SFGT-SN process are as follows:

Maximum particulate loading	Full loading ($\geq$10 gr/sft^3)
Pressure drop across the reactor	15-16 inches water
NH_3:NO_x mol ratio	(1.0-1.2):1
Average space velocity	5,000-8,000 ft^3/hr/ft^2
Temperature	750°F

Under these conditions both NO_x and SO_x removal efficiencies will be 90%.

Due to the unique processing scheme of the UOP SFGT-SN process, various additional benefits can be expected, particularly in the total energy consumption. Since the reactions occurring in the acceptance cycle are exothermic and the FGT system is located upstream of the air heater, the various heats of reaction are recoverable in the boiler air heater. The sensible heat of the regeneration off-gas is also recovered in the waste heat boiler and contact cooler. The process vendors also indicate that since SO_3 is removed by the sorbent, the dewpoint is lowered and therefore more sensible heat is potentially recoverable in the boiler air heater.

Shell International Petroleum Company developed this process and has been involved in the program since the early 1960's. The UOP Process Division holds the worldwide licensing rights to the process (with the exception of the Far East).

Analysis of Processing Subsections

The flow diagram and material balance for the base case are shown in Figure 8 and Table 37 respectively. The process is divided into six processing sections and the equipment assigned to the appropriate sections. The equipment list and descripions are shown in Table 38 by area. The total land requirement for this base case is estimated to be 4 acres.

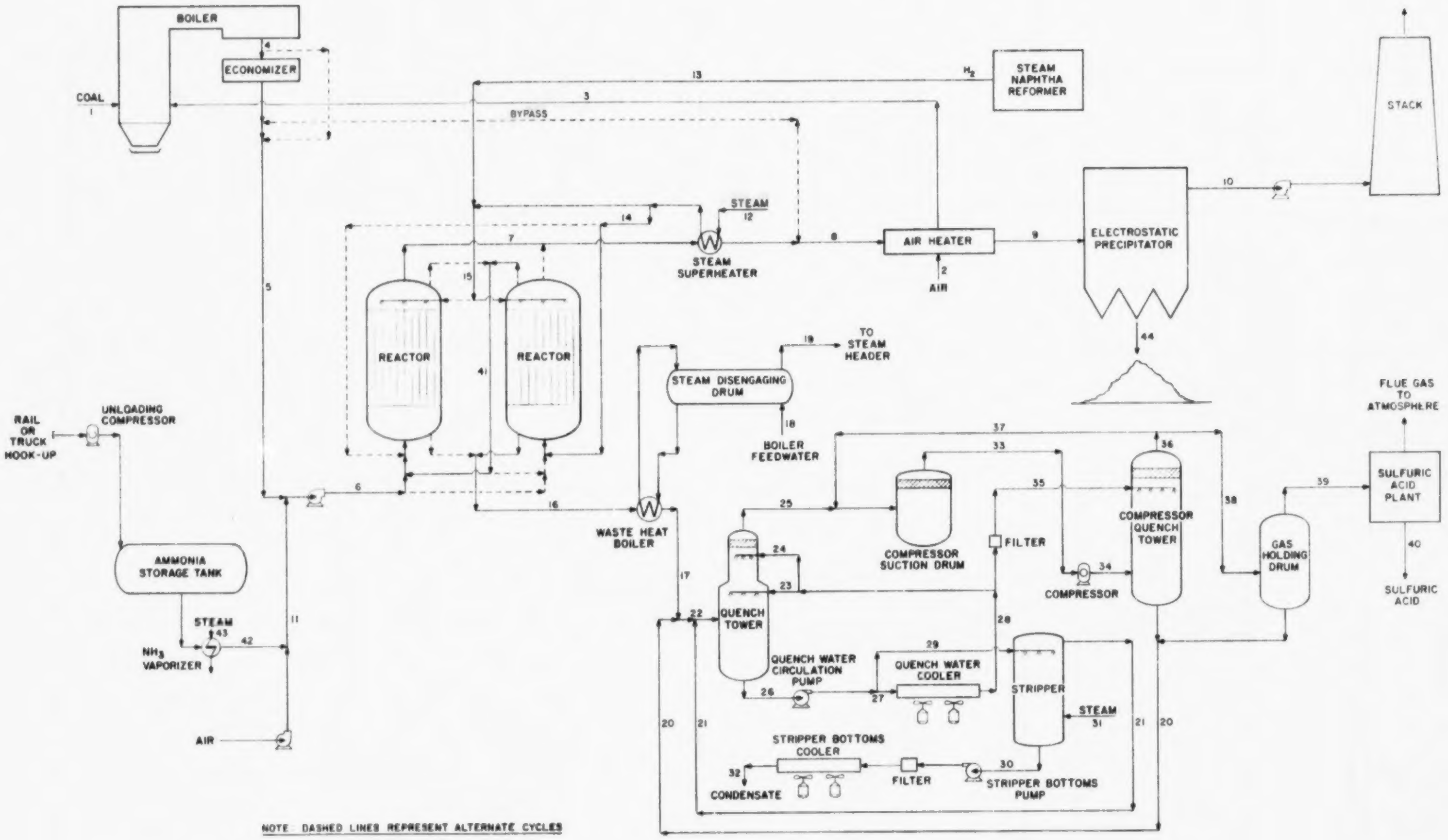
BOILER
ECONOMIZER
COAL
1
4
13
3
H2
BYPASS
STEAM NAPHTHA REFORMER
STACK
10
5
7
14
STEAM
12
STEAM SUPERHEATER
8
AIR HEATER
2
AIR
9
ELECTROSTATIC PRECIPITATOR
44
15
REACTOR
41
REACTOR
6
16
19
TO STEAM HEADER
STEAM DISENGAGING DRUM
18
BOILER FEEDWATER
RAIL OR TRUCK HOOK-UP
UNLOADING COMPRESSOR
AMMONIA STORAGE TANK
STEAM
43
42
11
NH_3 VAPORIZER
AIR
WASTE HEAT BOILER
17
22
25
24
23
QUENCH TOWER
26
QUENCH WATER CIRCULATION PUMP
27
QUENCH WATER COOLER
29
28
37
33
35
COMPRESSOR SUCTION DRUM
FILTER
34
COMPRESSOR
36
COMPRESSOR QUENCH TOWER
38
GAS HOLDING DRUM
39
SULFURIC ACID PLANT
FLUE GAS TO ATMOSPHERE
40
SULFURIC ACID
STRIPPER
STEAM
31
30
STRIPPER BOTTOMS PUMP
FILTER
STRIPPER BOTTOMS COOLER
32
CONDENSATE
20
21
21
20
NOTE: DASHED LINES REPRESENT ALTERNATE CYCLES

Figure 8. UOP SFGT-SN, SO_x-NO_x process. Flowsheet.

TABLE 37. UOP SFGT-SN, SO_x-NO_x PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to blower
1	Total stream, lb/hr	428,600	4,546,200	4,101,800	4,516,100	4,516,100
2						
3	Sft^3/min (60°F)		1,005,000	906,700	958,000	958,000
4	Temperature, °F		80	535	890	705
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	6	7	8	9	10
	Description	Flue gas-NH_3 mixture to reactor	Flue gas to steam superheater	Gas to air heater	Gas to ESP	Gas to stack
1	Total stream, lb/hr	4,576,700	4,548,800	4,548,800	4,993,200	4,938,600
2						
3	Sft^3/min (60°F)	971,800	969,900	969,900	1,068,100	1,068,100
4	Temperature, °F	720	798	786	300	300
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	11	12	13	14	15
	Description	NH_3 stream injected into flue gas	Steam to steam superheater	Hydrogen for regeneration	Steam to purge	Reducing gas to reactor
1	Total stream, lb/hr	60,600	65,500	2,870	6,270	62,100
2						
3	Sft^3/min (60°F)	13,800		5,700		26,200
4	Temperature, °F		298		650	
5	Pressure, psig		50		12	
6						
7						
8						
9						
10						

	Stream No.	16	17	18	19	20
	Description	Regeneration gas to waste heat boiler	Regeneration gas to quench tower	Boiler feedwater to steam disengaging drum	Steam to steam header	Compressor quench tower bottoms to quench tower
1	Total stream, lb/hr	96,250	96,250	12,710	12,710	28,520
2						
3	Sft^3/min (60°F)	28,100	28,100			
4	Temperature, °F			113	367	190
5	Pressure, psig			180	150	5
6						
7						
8						
9						
10						

(continued)

TABLE 37. (continued)

	Stream No.	21	22	23	24	25
	Description	Stripper overhead to quench tower	Quench tower feed	Lower quench tower recycle	Upper quench tower recycle	Quench tower overhead
1	Total stream, lb/hr	560	125,330	663,970	663,970	25,400
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F	190		113	113	113
5	Pressure, psig	4.5		125	125	5
6						
7						
8						
9						
10						

	Stream No.	26	27	28	29	30
	Description	Main quench tower bottoms	Main quench bottoms to cooler	Main quench bottoms from cooler	Main quench bottoms to stripper	Stripper bottoms to cooler
1	Total stream, lb/hr	1,427,870	1,355,570	1,355,570	72,300	76,590
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F	190	190	113	190	241
5	Pressure, psig		150	125	135	70
6						
7						
8						
9						
10						

	Stream No.	31	32	33	34	35
	Description	Steam to stripper	Stripper bottoms from cooler	Compressor suction	Compressor discharge	Compressor quench tower spray
1	Total stream, lb/hr	4,850	76,590	34,880	34,880	27,630
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F	298	113	122	356	113
5	Pressure, psig	50	65	4	65	125
6						
7						
8						
9						
10						

	Stream No.	36	37	38	39	40
	Description	Compressor quench overhead	Compressor spillback from quench	Gas holding drum feed	SO_2-rich compressed gas to sulfuric acid plant	Sulfuric acid product to storage
1	Total stream, lb/hr	33,990	9,480	24,510	24,510	35,600
2						
3	Sft^3/min (60^{o}F)					2,187
4	Temperature, ^{o}F	113	113	113	113	
5	Pressure, psig	65	4		30	
6						
7						
8						
9						
10						

(continued)

TABLE 37. (continued)

	Stream No.	41	42	43	44	
	Description	Purge gas to reactors in acceptance stage	NH_3 from storage	Steam to vaporizer	Fly ash from ESP	
1	Total stream, lb/hr	8,160	1,827	1,120	54,600	
2						
3	Sft^3/min (60°F)	2,900				
4	Temperature, °F		110	298		
5	Pressure, psig		225	50		
6						
7						
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TABLE 38. UOP SFGT-SN, SO_x-NO_x PROCESS

BASE CASE EQUIPMENT LIST
DESCRIPTION AND COST

Area 1--NH_3 Storage and Injection

Item	No.	Description	Total equipment cost, 1979 $
1. Compressor, NH_3 unloading	2	Single cylinder, double acting, 300 sft^3/min at 250 psig, 30 psig suction, 125 hp, cast iron	50,000
2. Tank, NH_3 storage	8	Horizontal, 9 ft dia x 66 ft long, 30,000 gal, 250 psig, carbon steel	230,100
3. Vaporizer, NH_3	2	Steam at 298°F, tube type, 30 ft^2, 1.03 MBtu/hr, carbon steel	36,800
4. Blower, NH_3 and air	2	7,000 aft^3/min, ΔP 5 in. H_2O, 10 hp, carbon steel	13,500
Subtotal			330,400

Area 2--Flue Gas Fans

Item	No.	Description	Total equipment cost, 1979 $
1. Blower, flue gas	2	1,088,400 aft^3/min, ΔP 17.1 in. H_2O, 4,000 hp, 316 stainless steel	1,510,300
Subtotal			1,510,300

(continued)

TABLE 38. (continued)

Area 3--Reactor Section

Item	No.	Description	Total equipment cost, 1979 $
1. Reactors	8	23.5 ft dia x 30 ft long, design temperatures - 1,060°F top, 800°F bottom, 22 psig internal, 1.37 psig external, carbon steel, insulated Included is a prorated allowance for a UOP-SFGT package consisting of reactor internals with initial catalyst fill, flue gas valves with hydraulic turning gear (2/reactor), process engineering, license fees, startup assistance, and for incremental air heater investment	16,969,000
2. Steam superheater	1	Steam at 298°F, tube type, 3,000 ft^2, 17.1 MBtu/hr, carbon steel	267,500
Subtotal			17,236,500

Area 4--Flow Smoothing

Item	No.	Description	Total equipment cost, 1979 $
1. Boiler, waste heat	1	Boiler feedwater at 113°F, 150 ft^2, 19.2 MBtu/hr, tube = 25 psig, shell = 200 psig, carbon steel	29,800
2. Steam disengaging drum	1	9 ft dia x 25 ft long, 420°F, 200 psig, carbon steel, insulated	38,100

(continued)

TABLE 38. (continued)

Item	No.	Description	Total equipment cost, 1979 $
3. Steam blowdown drum	1	Vertical, 5 ft dia x 15 ft long, 420°F, atmospheric pressure, carbon steel, insulated	13,700
4. Main quench tower	1	6 ft 6 in. top dia, 13 ft bottom dia, 40 ft long, 420°F, 50 psig, 317L stainless steel clad	189,600
5. Pump, quench water circulating	4	Centrifugal, 3,100 gpm, 368 ft head, 300 hp, casing, 317L stainless steel; impeller, 317L stainless steel	110,700
6. Filter, quench feed	1	65 ft^2 of cotton media, Dollinger model	30,000
7. Cooler, quench water	1	Air cooled fin fan type, 104.3 MBtu/hr, 23,600 ft^2, 110 psig, 317L stainless steel	793,100
8. Stripper	1	6 ft dia x 20 ft long, 280°F, 50 psig, 317L stainless steel clad	51,100
9. Pump, stripper bottoms	2	Centrifugal, 155 gpm, 172 ft head, 15 hp, casing, 317L stainless steel; impeller, 317L stainless steel	16,000
10. Filter, stripper bottoms	1	38 ft^2 of cotton media, Dollinger model	15,000
11. Cooler, stripper bottoms	1	Air cooled fin fan type, 9.96 MBtu/hr, 1,800 ft^2, 90 psig, carbon steel	50,000

(continued)

TABLE 38. (continued)

Item	No.	Description	Total equipment cost, 1979 $
12. Compressor suction drum	1	5 ft dia x 10 ft long, 175°F, 50 psig, 317L stainless steel clad	41,800
13. Compressor, regeneration rich gas	1	4,300 aft^3/min at 122°F, suction temperature, 122°F; suction pressure, 967 mm Hg abs; discharge pressure, 4,121 mm Hg abs, 550 hp, 900 rpm (max), 317L stainless steel	100,000
14. Compressor, quench tower	1	4 ft dia x 14 ft long, 410°F, 90 psig, 317L stainless steel clad, insulated	68,100
15. Gas holding drum	1	12 ft dia x 24 ft long, 410°F, 90 psig, 317L stainless steel clad, insulated	127,000
Subtotal			1,674,000

Area 5--Steam Naphtha Reformer

Item	No.	Description	Total equipment cost, 1979 $
1. Packaged steam naphtha reformer	1	10 Msft^3/day, 1,500 gal/hr naphtha, plot area = 50 ft x 75 ft, skid mounted	3,222,000
Subtotal			3,222,000

(continued)

TABLE 38. (continued)

Area 6--H_2SO_4 Plant

	Item	No.	Description	Total equipment cost, 1979 $
1.	H_2SO_4 plant, contact unit	1	427 tons/day, 100% H_2SO_4 as 98% or 93% acid, flue gas meets EPA standards	5,719,000
	Subtotal			5,719,000
	Total, Areas 1-6			29,692,200

NH3 Storage and Injection--

A compressor (including a spare) for unloading liquid NH_3 received by truck or rail and 250 psig storage tanks for a 30-day supply of NH_3 are included in this area. A steam-heated, shell-and-tube vaporizer is needed for vaporization of NH_3, before it is mixed with air. Air is supplied by a blower to form a 5% NH_3-in-air mixture for better dispersion in the flue gas. The blower is sized based on a 5-inch water pressure drop. The NH_3 injection grid is primarily piping and the cost is incorporated into the area investment estimate shown in the summary capital investment estimates in the following report section, Economic Evaluation and Comparison.

Gas Handling--

Two FD blowers are provided between the boiler economizer and the reactors to overcome the pressure drop of the reactor system. A pressure drop of 17.1 inches water for the system is used as a basis. Since these fans are booster fans for the NO_x removal system, the total investment and revenue requirements are included in the FGT costs. Other fan requirements for boiler and ESP units are not included.

Reactor Section--

Eight reactors are used, six of which are on stream while two are in the regeneration stage. The reactors are parallel-passage design with the catalyst contained in unit cells. The flue gas is passed over the catalyst surface rather than through it. Reactor operating sequences are controlled by 16 large flue gas valves, a hydraulic valve switching system and a sequence controller. A steam superheater between the reactors and air heater heats steam for purging during catalyst regeneration. Also included in this section is the incremental investment required for a larger air heater to recover portions of the claimed heat credits.

Flow Smoothing--

The off-gas from regeneration requires treatment for heat recovery, water condensation to produce an SO_2-rich stream, and dampening of cyclic flow. A waste heat boiler, steam disengaging drum, and steam blowdown drum are used in the transfer of heat from the regeneration off-gas to boiler feedwater. Quench towers and strippers are used in water condensation and the production of a concentrated SO_2 gas stream. A gas compressor and gas holder are used in dampening the flow of the SO_2-rich gas to the sulfuric acid manufacturing plant.

Steam-Naphtha Reformer--

Hydrogen gas for regeneration of the reactors is manufactured in a steam-naphtha reformer producing about 10 Msft3/day of 95% purity hydrogen. The naphtha requirement is 1,500 gal/hr (assuming 80% for feed and 20% for fuel). The electricity required is 250 kW and 12.4 klb/hr of steam is produced for other uses.

Sulfuric Acid Plant--

There are several possible uses of the SO_2-rich off-gas from the regeneration process, i.e., conversion to liquefied SO_2, elemental sulfur, or H_2SO_4. In this study, on the process developer recommendation, H_2SO_4 is produced. A contact unit producing 427 tons/day of 100% H_2SO_4 as 98% acid is used. The electricity required is 2,050 kW and the byproduct steam available for other uses is 32 Mlb/hr. The flue gas from the sulfuric acid plant meets EPA standards. Savings could be realized if a less expensive acid plant were used in which the flue gas from the acid plant would be combined with the boiler flue gas for SO_x-NO_x FGT.

LIMESTONE FGD

The limestone FGD system with four trains of absorbers is designed for 90% removal of SO_2. The system does not include an ESP or a prescrubber. Each of the absorbers is a TCA type with a presaturator for cooling and humidifying the flue gas. Each scrubber is equipped with a chevron-type entrainment separator at the scrubber outlet. Stack gas reheat to 175°F for plume buoyancy is provided by indirect steam reheat of the cleaned gas. Waste sludge is disposed of by ponding as a 15 weight percent solids slurry. The settled sludge contains 60% free water; the excess water is returned to the scrubber system.

A complete process description, flowsheet, material balance, and equipment list for the limestone FGD process used in this study may be found in the following EPA report, Definitive SO_x Control Process Evaluations: Limestone, Double Alkali, and Citrate FGD Processes, by Tomlinson, S. V., et al (17).

ESP

Four ESP's of varying efficiencies are included in this study to allow overall SO_x-NO_x-PM control system comparability. Four 99.5% efficient cold ESP's are used with the Hitachi Zosen and UOP SFGT-N processes. Each ESP has about 117,000 ft^2 of plate area and handles about 391,000 aft^3/min of flue gas at 300°F. The estimated total equipment cost in 1979 dollars is $6,141,000.

Four larger 99.5% efficient cold ESP's are used with the UOP SFGT-SN process since this process also removes some SO_3, thus increasing the difficulty for fly ash removal. Each ESP has about 180,000 ft^2 of plate area and handles about 390,000 aft^3/min of flue gas at 300°F. The estimated total equipment cost in 1979 dollars is $7,934,000.

Four 98% efficient hot ESP's are used with the Kurabo Knorca process. The remaining 1.5% PM is removed downstream at the NO_x reactor. Each ESP has 129,000 ft^2 of plate area and treats about 537,000 aft^3/min of flue gas at 705°F. The estimated total equipment cost in 1979 dollars is $6,924,000.

One 94% efficient cold ESP is used with the Moretana Calcium process while the remaining PM is removed in the downstream prescrubber. This ESP has 232,000 ft^2 of collecting area and handles about 1,544,000 aft^3/min of flue gas at 300°F. The estimated total equipment cost in 1979 dollars is $4,429,000.

ECONOMIC EVALUATION AND COMPARISON

Based on the definitive power plant, process design, and economic premises outlined earlier in this report and on the specific process equipment for each process described in the previous section, total capital investment and total annual revenue requirements were prepared for the economic evaluation and comparison of the seven alternative NO_x FGT systems. It should be emphasized once again that in some cases, particularly the dry NO_x-only processes, additional costs for FGD and ESP must be added to the capital investment and annual revenue requirement for NO_x-only removal to provide an equal comparison between the NO_x FGT alternatives.

PROCEDURE FOR ESTIMATING THE TOTAL CAPITAL INVESTMENTS

From material balance calculations and information supplied by the process vendors, the various pieces of equipment for each process were sized and costed. The process equipment costs were obtained primarily by using in-house cost data but vendor quotes were used for unique or unusual equipment. This cost for each piece of equipment reflected the midwestern power plant location and was escalated to mid-1979 dollars by assuming an escalation rate of 8%/yr. Tables containing the equipment list, equipment description, and the mid-1979 equipment cost (material and labor) for each process are included in the Systems Estimated section.

Because of the preliminary nature of this study, the various installation expenses such as piping, foundations, etc., were estimated as a percentage of equipment costs based on both accepted ranges from published data and previous TVA experience. The subtotal direct investment is the sum of the process equipment costs and installation expenses. The final component of the total direct investment is the services and miscellaneous cost which were calculated as percentages of the subtotal direct investment. The sum of the subtotal direct investment plus the services and miscellaneous cost is the total direct investment.

The total indirect investment (the sum of various indirect investment, such as engineering design and supervision, A&E contractor, construction expense, and contractor fees) was calculated based on the equipment costs or the total direct investment using the formulas described in the Economic Premises section. A contingency calculated as 20% of the sum of the total direct and total indirect investment was also included. The sum of the total direct investment, the total indirect investment, and the contingency is the total fixed investment.

Other capital charges consist of an allowance for startup and modification, interest during construction, land costs, working capital, and royalty fee. The allowance for startup and modification and interest during construction were calculated as a percentage of the total fixed investment. The total depreciable investment is the sum of the total fixed investment, the previously calculated allowance for startup and modification, and interest during construction. The sum of total depreciable investment, land costs, working capital, and royalty fee is the total capital investment.

RESULTS

The total capital investment for each of the seven NO_x FGT alternatives calculated for a base case, new, 500-MW coal-fired power plant is listed on the following pages.

Asahi SO_x-NO_x Process

Total Capital Investment--

The total capital investment for the base case application for the Asahi process was estimated at $104.9M ($233/kW) in mid-1979 dollars. This total cost can be broken down into the various investment cost categories shown in Table 39. The direct investment accounts for 54.1% of the estimated total capital investment for the Asahi process. The indirect investments make up 24.5% of the required investment and the various other capital charges make up the remaining 21.4%.

The direct investment can be further broken down into costs by processing area also shown in Table 39. The cracking, SO_x-NO_x absorption, and crystallization areas alone account for about 62% of the total direct investment. The high costs of these areas are due to the use of a liquid scrubbing system and the resulting necessity of investment for large reactors, evaporators, crystallizers, a dryer, rotary kiln crackers, and screw decanter separators. Although the absorption section consists of four trains with only four pieces of equipment in each train, the direct investment for this section is nearly $11.5M, reflecting the difficulty in absorbing low concentrations of NO_x into aqueous solution. This high investment cost for absorbers must be, however, compared with an expensive generating system to produce a gas-phase oxidant in the oxidation processes.

Major indirect investments were the construction expense at $7.1M, engineering design and supervision at about $2.2M, and contractors fees at $2.1M. A&E contractors expense was estimated at $0.5M. A 20% contingency of $13.8M was assessed.

The remaining $22.4M of the total capital investment was made up of other capital charges, such as allowance for startup and modification, interest during construction, land, working capital, and royalty fee. Ten percent of the total fixed investment ($8.3M) was assessed as an allowance

TABLE 39. ASAHI SO_x-NO_x PROCESS

SUMMARY OF ESTIMATED TOTAL CAPITAL INVESTMENT[a]

(500-MW new coal-fired power unit,
3.5% S in coal, 90% NO_x removal, 99% SO_2 removal)

	Investment, $	% of total direct investment
Direct investment[b]		
Materials handling (hoppers, conveyors, and dust collector)	1,013,000	1.8
Feed preparation (crusher, ball mill, tanks, agitators, and pumps)	1,288,000	2.3
Gas handling (booster fans and gas ducts and dampers from absorbers to stack)	5,105,000	9.0
Particulate control (common feed plenum and gas ducts and dampers from plenum to prescrubber, four prescrubbers, thickener, tanks, agitators, and pumps)	2,617,000	4.6
SO_x-NO_x absorption (four absorbers, tanks, agitators, and pumps)	11,407,000	20.1
Crystallization (evaporators, crystallizers, refrigeration units, screw decanters, tanks, agitators, and pumps)	11,395,000	20.0
Nitrogen treatment (screw decanter, cracker, filter, conveyors, tanks, agitators, and pumps)	1,168,000	2.0
Cracking (dryer, crackers, furnace, blowers, conveyors, tanks, and pumps)	12,328,000	21.7
Gypsum production (thickener, screw decanter, conveyor, tanks, agitators, and pumps)	2,385,000	4.2
Absorbent regeneration (stripper, thickeners, conveyors, tanks, agitators, and pumps)	3,684,000	6.5
Raw material storage (hoppers, conveyors, tanks, agitator, and pumps)	850,000	1.5
Solids disposal	46,000	0.1
Subtotal	53,286,000	93.8
Services, utilities, and miscellaneous	3,197,000	5.6
Trucks and earthmoving equipment	328,000	0.6
Total direct investment	56,811,000	100.0
Indirect Investment		
Engineering design and supervision	2,181,000	3.8
Architect and engineering contractor	545,000	1.0
Construction expense	7,147,000	12.6
Contractor fees	2,068,000	3.6
Total indirect investment	11,941,000	21.0
Contingency	13,750,000	24.2
Total fixed investment	82,502,000	145.2
Other Capital Charges		
Allowance for startup and modification	8,250,000	14.5
Interest during construction	9,900,000	17.5
Total depreciable investment	100,652,000	177.2
Land	254,000	0.5
Working capital	3,703,000	6.5
Royalty fee	300,000	0.5
Total capital investment	104,909,000	184.7

$/aft^3/min	$/kW
48.9	233

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a. Basis

Midwest plant location represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979.

Stack gas reheat to 175°F by indirect steam reheat.

Minimum in-process storage; only pumps are spared.

Disposal pond located 1 mile from power plant.

Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.

Construction labor shortages with accompanying overtime pay incentive not considered.

b. Each area direct investment includes total equipment costs (given in the equipment lists) plus installation labor and material costs for electrical, piping, ductwork, foundations, structural, instrumentation, insulation, and site preparation.

for plant startup and modifications. Interest during construction was estimated as 12.0% of the total fixed investment, or \$9.9M. The total cost for land was estimated at about \$0.25M and working capital at \$3.7M. A royalty fee of \$0.3M was also added.

IHI SO_x-NO_x Process

Total Capital Investment--

The estimated total capital investment for the IHI SO_x-NO_x process was \$203.6M (\$482/kW) in mid-1979 dollars as shown in Table 40. The direct investment accounts for 57.6% of the total capital investment. The indirect investments make up 22.5% of the total investment and the various other capital charges make up the remaining 19.9%.

The direct investment for the IHI process can be further subdivided into the various processing areas as shown in Table 40. The direct investment for the ozone generating system dwarfs the investments required for the other processing areas, contributing 66%, about two-thirds of the total direct investment and 38.1% of the total capital investment. Other capital-intensive processing areas include: particulate control, \$10.0M; SO_x-NO_x absorption, \$7.2M; and nitrate treatment, \$5.2M.

Major indirect investments were the construction expense, \$13.1M and contractors fees at \$3.6M. Engineering design and supervision and A&E contractor expense contributed significantly less at \$1.5M and \$0.4M respectively. The remaining indirect investment, project contingency, was \$27.2M.

The remaining \$40.5M of the total capital investment was made up of other capital charges including: allowance for startup and modifications, interest during construction, land, working capital, and royalty fee. The allowance for startup and modifications and interest during construction were assessed at 10% (\$16.3M), and 12% (\$19.6M), respectively, and contributed most of the other capital charges. Based on information from the process vendors on land requirements and from material balance calculations, land and working capital were estimated as \$0.29M and \$4.0M. The royalty fee added was \$0.3M.

Moretana Calcium SO_x-NO_x Process

Total Capital Investment--

The total capital investment of the Moretana Calcium process was \$88.0M (\$190/kW). This can be broken down into the direct investment, the indirect investment, and other capital charges as shown in Table 41. The direct investment at slightly more than \$46.7M accounts for 53.1% of the total capital investment. The indirect investments are 24.5% of the toal capital investment, while other capital charges constitute the remaining 22.4%.

The direct investment for the Moretana Calcium process can be further subdivided into the various processing areas shown in Table 41. The major processing area, in terms of the direct investment involved, is the chlorine dioxide generating system. A direct investment of nearly \$16.8M, 19.1% of the

TABLE 40. IHI SO_x-NO_x PROCESS

SUMMARY OF ESTIMATED TOTAL CAPITAL INVESTMENT[a]

(500-MW new coal-fired power unit,
3.5% S in coal, 90% NO_x removal, 90% SO_2 removal)

	Investment, $	% of total direct investment
Direct Investment[b]		
Material handling (hoppers, conveyors, and dust collector)	1,013,000	0.8
Feed preparation (crusher, ball mill, tanks, agitators, and pumps)	1,153,000	1.0
Gas handling (common plenum, ducts and dampers from plenum to prescrubber, and booster fans)	4,371,000	3.7
Particulate control (prescrubber, thickeners, filter, tanks, agitators, and pumps)	9,952,000	8.5
SO_x-NO_x absorption (absorber and pumps)	7,239,000	6.2
Stack gas reheat (single indirect steam reheater)	1,016,000	0.9
Gypsum production (oxidation tower, thickener, centrifuge, tanks, agitators, and pumps)	2,546,000	2.2
Nitrate treatment (evaporators, thickener, decomposer, blowers, conveyors, furnace, tanks, agitators, and pumps)	5,153,000	4.4
Raw material storage (silo, conveyors, tanks, agitators, and pumps)	328,000	0.3
Solids disposal	50,000	-
Ozone generating system (one unit complete)	77,562,000	66.1
Subtotal	110,383,000	94.1
Services, utilities, and miscellaneous	6,623,000	5.6
Trucks and earthmoving equipment	364,000	0.3
Total direct investment	117,370,000	100.0
Indirect Investment		
Engineering design and supervision	1,540,000	1.3
Architect and engineering contractor	385,000	0.3
Construction expense	13,052,000	11.1
Contractor fees	3,590,000	3.1
Total indirect investment	18,567,000	15.8
Contingency	27,187,000	23.2
Total fixed investment	163,124,000	139.0
Other Capital Charges		
Allowance for startup and modification	16,312,000	13.9
Interest during construction	19,575,000	16.7
Total depreciable investment	199,011,000	169.6
Land	285,000	0.2
Working capital	4,026,000	3.4
Royalty fee	300,000	0.3
Total capital investment	203,622,000	173.5
	$/aft^3/min	$/kW
	94.9	482

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a. Basis
Midwest plant location represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979.
Stack gas reheat to 175°F by indirect steam reheat.
Minimum in-process storage; only pumps are spared.
Disposal pond located 1 mile from power plant.
Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.
Construction labor shortages with accompanying overtime pay incentive not considered.

b. Each area direct investment includes total equipment costs (given in the process equipment list) plus installation labor and material costs for electrical, piping, ductwork, foundations, structural, instrumentation, insulation, and site preparation.

TABLE 41. MORETANA CALCIUM PROCESS

SUMMARY OF ESTIMATED TOTAL CAPITAL INVESTMENT[a]

(500-MW new coal-fired power unit,
3.5% S in coal, 90% NO_x removal, 95% SO2 removal)

	Investment, $	% of total direct investment
Direct Investment[b]		
Material handling (hoppers, conveyors, and dust collector)	1,111,000	2.4
Feed preparation (crusher, ball mill, tanks, agitators, and pumps)	1,461,000	3.1
Gas handling (common plenum, ducts and dampers from plenum to prescrubber, and booster fans)	4,246,000	9.1
Particulate control (two prescrubbers, tanks, agitators, and pumps)	1,591,000	3.4
SO_x-NO_x absorption (two absorbers, clarifier, tanks, agitators, and pumps)	4,640,000	9.9
Stack gas reheat (two indirect steam reheaters)	1,359,000	2.9
Gypsum production (centrifuge, conveyor, tanks, agitators, and pumps)	2,732,000	5.8
Nitrogen treatment (centrifuge, evaporators, conveyor, tanks, agitators, and pumps)	2,618,000	5.6
Raw material storage (silos, hoppers, and conveyors)	918,000	2.0
Solids disposal (sludge pond, tanks, agitators, and pumps)	6,563,000	14.1
ClO_2 generation (two skid-mounted units complete)	16,809,000	36.0
Subtotal	44,048,000	94.3
Services, utilities, and miscellaneous	2,643,000	5.7
Total direct investment	46,691,000	100.0
Indirect Investment		
Engineering design and supervision	1,955,000	4.2
Architect and engineering contractor	440,000	1.0
Construction expense	5,975,000	12.8
Contractor fees	1,782,000	3.8
Total indirect investment	10,161,000	21.8
Contingency	11,370,000	24.3
Total fixed investment	68,222,000	146.1
Other Capital Charges		
Allowance for startup and modification	6,166,000	13.2
Interest during construction	8,187,000	17.5
Total depreciable investment	82,575,000	176.8
Land	910,000	2.0
Working capital	4,182,000	9.0
Royalty fee	300,000	0.6
Total capital investment	87,967,000	188.4

$/aft^3/min	$/kW
41.0	190

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a. Basis
Midwest plant location represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979.
Stack gas reheat to 175°F by indirect steam reheat.
Minimum in-process storage; only pumps are spared.
Disposal pond located 1 mile from power plant.
Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.
Construction labor shortages with accompanying overtime pay incentive not considered.

b. Each area direct investment includes total equipment costs (given in the process equipment list) plus installation labor and material costs for electrical, piping, ductwork, foundations, structural, instrumentation, insulation, and site preparation.

total capital investment, is required. Other capital-intensive processing areas are SO_x-NO_x absorption, \$4.6M and solids disposal, \$6.6M.

Major indirect investments are construction expense, \$6.0M; engineering design and supervision, \$2.0M; and contractors fees, \$1.8M. A&E contractor expense, \$0.50M, was significantly less. The project contingency was \$11.4M.

The allowance for startup and modifications and interest during construction were assessed at 10% (\$6.2M), and 12% (\$8.2M), respectively, of the total fixed investment. These two capital charges together contributed nearly 16.3% of the total capital investment and made up most of the other capital charges. Based on information from the process vendor on land requirements and from material balance calculations, land and working capital were estimated at \$0.9M and \$4.2M. A \$0.3M royalty fee was added.

Hitachi Zosen NO_x-Only Process

The total capital investment for the Hitachi Zosen NO_x-only process was estimated to be \$23.3M (\$48.2/kW). Table 42 shows a breakdown of this total capital investment. The direct investment composes 53.4% of the estimated total capital investment. The indirect investment constitutes 26.2% of the total capital investment. The other capital charges compose the remaining 20.4%.

The direct investment comprises the three major processing area costs shown in Table 42. The cost for the reactor section accounts for 83.7% of the total direct investment. This section, as for most of the other dry NO_x removal processes, is the heart of the process.

The major indirect investments were the construction expense, \$2.0M and contractor fees, \$0.7M. Both were based on the total direct investment. The engineering design and supervision at \$0.3M and the A&E contractor expense at \$0.07M were based on the number of pieces of equipment. A contingency of \$3.1M was also assessed.

Under other capital charges, allowance for startup and modifications was \$1.9M which is 10% of the total fixed investment. Interest during construction was estimated as 12.0% of the total fixed investment of \$2.2M. Land cost was estimated at \$5,000 while working capital was estimated at \$0.4M. A \$0.3M royalty fee, as stated by the vendor, was added.

Kurabo Knorca NO_x-Only Process

The total capital investment of the Kurabo Knorca process was estimated to be \$21.2M (\$43.9/kW). This total capital investment is broken down into direct investment, indirect investment, contingency, and other capital charges as shown in Table 43. The direct investment composes 51.4% of the total capital investment, or slightly less than \$10.9M. The indirect investment and contingency account for 28.2% of the total capital investment, or about \$6.0M. The other capital charges make up the remaining 20.4%.

TABLE 42. HITACHI ZOSEN PROCESS

SUMMARY OF ESTIMATED CAPITAL INVESTMENT[a]

(500-MW new coal-fired power unit, 3.5% S in coal; 90% NO_x removal)

	Investment, $	% of total direct investment
Direct Investment[b]		
NH_3 storage and injection (compressors, tanks, vaporizers, and blowers)	688,000	5.5
Reactor section (reactors, catalyst fill)	10,419,000	83.7
Gas handling (flue gas fans)	635,000	5.1
Subtotal	11,742,000	94.3
Services, utilities, and miscellaneous	705,000	5.7
Total direct investment	12,447,000	100.0
Indirect Investment		
Engineering design and supervision	274,000	2.2
Architect and engineering contractor	69,000	0.6
Construction expense	2,027,000	16.3
Contractor fees	652,000	5.2
Total indirect investment	3,022,000	24.3
Contingency	3,094,000	24.8
Total fixed investment	18,563,000	149.1
Other Capital Charges		
Allowance for startup and modifications	1,856,000	14.9
Interest during construction	2,228,000	17.9
Total depreciable investment	22,647,000	181.9
Land	5,000	0.0
Working capital	355,000	2.9
Royalty fee	300,000	2.4
Total capital investment	23,307,000	187.2

$/aft^3/min	$/kW
10.9	48.2

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a. Basis
Midwest plant location represents project beginning mid-1977, ending mid-1980. Average basis for scaling, mid-1979.
Investment requirements for fly ash disposal excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

b. Each area direct investment includes total equipment costs (given in the process equipment list) plus installation labor and material costs for electrical, piping, ductwork, foundations, structural, instrumentation, insulation, and site preparation.

TABLE 43. KURABO KNORCA PROCESS

SUMMARY OF ESTIMATED CAPITAL INVESTMENT[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
90% NOx removal)

	Investment, $	% of total direct investment
Direct Investment[b]		
NH_3 storage and injection (compressors, tanks, and vaporizers)	691,000	6.3
Reactor section (static mixer, reactor, feeder, conveyors, hopper, elevator, and catalyst fill)	8,898,000	81.7
Gas handling (flue gas fans)	635,000	5.8
Waste disposal (filters and blowers)	58,000	0.5
Subtotal	10,282,000	94.3
Services, utilities, and miscellaneous	617,000	5.7
Total direct investment	10,899,000	100.0
Indirect Investment		
Engineering design and supervision	599,000	5.5
Architect and engineering contractor	150,000	1.4
Construction expense	1,815,000	16.6
Contractor fees	590,000	5.4
Total indirect investment	3,154,000	28.9
Contingency	2,811,000	25.8
Total fixed investment	16,864,000	154.7
Other Capital Charges		
Allowance for startup and modifications	1,686,000	15.5
Interest during construction	2,024,000	18.6
Total depreciable investment	20,574,000	188.8
Land	5,000	0.0
Working capital	311,000	2.9
Royalty fee	300,000	2.7
Total capital investment	21,190,000	194.4

$/aft^3/min	$/kW
9.87	43.9

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a. Basis
Midwest plant location represents project beginning mid-1977, ending mid-1980. Average basis for scaling, mid-1979.
Investment requirements for fly ash disposal excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

b. Each area direct investment includes total equipment costs (given in the process equipment list) plus installation labor and material costs for electrical, piping, ductwork, foundations, structural, instrumentation, insulation, and site preparation.

The direct investment for the Kurabo Knorca process may be subdivided as shown in Table 43. The reactor section, which contains the reactors, catalyst, static gas mixer, catalyst feeders and hoppers, conveyors, and bucket elevators, is the most costly area at almost \$8.9M. This represents 81.7% of the total direct investment.

The construction expense, the major indirect investment of \$1.8M, and contractor fees of \$0.6M were based on the total direct investment. The engineering design and supervision, estimated at about \$0.6M, and A&E contractor expense, estimated at \$0.2M, were based on the number of pieces of equipment. The contingency was \$2.8M.

The other capital charges totaled about \$4.3M. Included were allowance for startup and modification, \$1.7M, and interest during construction, \$2.0M, which were estimated at 10% and 12%, respectively, of the total fixed investment. Land and working capital were \$5,000 and \$0.3M respectively. A royalty fee of \$0.3M was also added.

UOP SFGT-N, NO_x-Only Process

The total capital investment of the UOP SFGT-N process was estimated to be about \$18.4M (\$38.1/kW). Table 44 shows the subdivision of this total capital investment. The direct investment represents 53.4% of the total capital investment, the indirect investment and contingency are 27.0% and other capital charges are the remaining 19.6%.

The direct investment comprises the three major processing area costs shown in Table 44. The cost for the reactor section accounts for 80.9% of the total direct investment. This section contains the reactors, unit cells, and initial catalyst fill.

The major indirect investments were the construction expense of about \$1.7M and contractor fees of \$0.5M. The engineering design and supervision of \$0.2M and the A&E contractor expense of \$0.06M were derived from the number of pieces of equipment. The contingency was \$2.5M.

Other capital charges include allowance for startup and modification, interest during construction, land, and working capital. Allowance for startup and modification was taken as 10% of the total fixed investment, or about \$1.5M. Interest during construction was estimated as 12% of the total fixed investment, or \$1.8M. Land and working capital were estimated to be \$5,000 and \$0.3M respectively. The royalty fee was prorated and included under the direct investment.

UOP SFGT-SN, SO_x-NO_x Process

The total capital investment for the UOP SFGT-SN process was estimated to be \$67.2M (\$139/kW). Table 45 shows a breakdown of this total capital investment. The direct investment and indirect investment plus contingency account for 55.6% and 24.4%, respectively, of the total capital investment. The remaining 20% comprises the other capital charges.

TABLE 44. UOP SFGT-N, NO_x-ONLY PROCESS

SUMMARY OF ESTIMATED CAPITAL INVESTMENT[a]

(500-MW new coal-fired power unit, 3.5% S in coal; 90% NO_x removal)

	Investment, $	% of total direct investment
Direct Investment[b]		
NH_3 storage and injection (compressors, tanks, vaporizers, and blowers)	688,000	7.0
Reactor section (reactors, catalyst fill)	7,964,000	80.9
Gas handling (flue gas fans)	635,000	6.4
Subtotal	9,287,000	94.3
Services, utilities, and miscellaneous	557,000	5.7
Total direct investment	9,844,000	100.0
Indirect Investment		
Engineering design and supervision	236,000	2.4
Architect and engineering contractor	59,000	0.6
Construction expense	668,000	16.9
Contractor fees	546,000	5.6
Total indirect investment	2,509,000	25.5
Contingency	2,471,000	25.1
Total fixed investment	14,824,000	150.6
Other Capital Charges		
Allowance for startup and modifications	1,482,000	15.0
Interest during construction	1,779,000	18.1
Total depreciable investment	18,085,000	183.7
Land	5,000	0.0
Working capital	333,000	3.4
Total capital investment	18,423,000	187.1

$/aft^3/min	$/kW
8.58	38.1

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a. Basis
Midwest plant location represents project beginning mid-1977, ending mid-1980. Average basis for scaling, mid-1979.
Investment requirements for fly ash disposal excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

b. Each area direct investment includes total equipment costs (given in the process equipment list) plus installation labor and material costs for electrical, piping, ductwork, foundations, structural, instrumentation, insulation, and site preparation.

TABLE 45. UOP SFGT-SN, SO_x-NO_x PROCESS

SUMMARY OF ESTIMATED CAPITAL INVESTMENT[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
90% NO_x removal, 90% SO_2 removal)

	Investment, $	% of total direct investment
Direct Investment[b]		
NH_3 storage and injection (compressors, tanks, vaporizers, and blowers)	688,000	1.8
H_2SO_4 plant (contact unit)	7,172,000	19.2
Reactor section (reactors, catalyst fill, superheater, valves, hydraulic valve switching system, unit cells)	19,110,000	51.2
Flow smoothing section (boiler, drums, quench towers, pumps, filters, strippers, fans, and compressors)	2,545,000	6.8
Steam naphtha reformer (one reformer unit)	4,105,000	11.0
Gas handling (flue gas fans)	1,589,000	4.3
Subtotal	35,209,000	94.3
Services, utilities, and miscellaneous	2,113,000	5.7
Total direct investment	37,322,000	100.0
Indirect Investment		
Engineering design and supervision	709,000	1.9
Architect and engineering contractor	177,000	0.5
Construction expense	5,043,000	13.5
Contractor fees	1,503,000	4.0
Total indirect investment	7,432,000	19.9
Contingency	8,951,000	24.0
Total fixed investment	53,705,000	143.9
Other Capital Charges		
Allowance for startup and modifications	5,371,000	14.4
Interest during construction	6,445,000	17.3
Total depreciable investment	65,521,000	175.6
Land	14,000	0.0
Working capital	1,618,000	4.3
Total capital investment	67,153,000	179.9

$/aft³/min	$/kW
31.3	139

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a. Basis
Midwest plant location represents project beginning mid-1977, ending mid-1980. Average basis for scaling, mid-1979.
Investment requirements for fly ash disposal excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

b. Each area direct investment includes total equipment costs (given in the process equipment list) plus installation labor and material costs for electrical, piping, ductwork, foundations, structural, instrumentation, insulation, and site preparation.

The direct investment is to be subdivided into the cost areas as shown in Table 45. The reactor section, containing reactors, unit cells, initial catalyst fill, steam superheater, hydraulic valve switching system and valves, is the single largest direct investment factor, i.e., \$19.1M or 51.2% of the total direct investment. The sulfuric acid plant, steam naphtha reformer, and flow smoothing sections are also capital-intensive process areas.

Major indirect investments were the construction expense at about \$5.0M and contractor fees at \$1.5M. Engineering design and supervision and A&E contractor expenses were about \$0.7M and \$0.2M respectively. Contingency was estimated to be \$9.0M.

The other capital charges were approximately \$5.4M for allowance for startup and modifications, \$6.4M for interest during construction, \$14k for land, and \$1.6M for working capital. The royalty fee was prorated and included with the direct investment.

Limestone FGD

The total capital investment of the limestone FGD process with 90% SO_2 removal, which is used in addition to each of the dry NO_x-only FGT processes for overall system comparability with SO_x-NO_x processes, was estimated to be \$50.4M (17).

ESP

The total capital investment of the various ESP's was estimated using the same procedures as previously given for the NO_x processes. These investments are summarized below.

1. The total capital investment, in 1979 dollars, for four 99.5% efficient cold ESP's, to be used with the Hitachi Zosen, and UOP SFGT-N processes, was estimated to be \$10,830,000. This was based on a total equipment cost of \$6,141,000.

2. The total capital investment, in 1979 dollars, for four 99.5% efficient cold ESP's, to be used with the UOP SFGT-SN process, was estimated to be \$14,625,000. This was based on a total equipment cost of \$7,934,000.

3. The total capital investment, in 1979 dollars, for four 98% efficient hot ESP's, to be used with the Moretana Calcium process, was estimated to be \$12,146,000. This was based on a total equipment cost of \$6,924,000.

4. The total capital investment, in 1979 dollars, for one 94% efficient cold ESP, to be used with the Moretana Calcium process, was estimated to be \$7,213,000. This was based on a total equipment cost of \$4,429,000.

PROCEDURE FOR ESTIMATING THE AVERAGE ANNUAL REVENUE REQUIREMENTS

The annual revenue requirements were calculated based on a midwestern power plant location and mid-1980 costs. These average annual revenue requirements are divided into the direct costs, which consist of raw materials costs and conversion costs and indirect costs, which consist of capital charges and overheads.

The raw material costs were calculated by determining the current (1978) delivered price of these raw materials in the Chicago area and assuming an escalation rate of 8%/yr for the delivered raw material. For the conversion costs, the utility and labor costs were projected using internal TVA information. Operating labor and supervision were based on both process vendor information and previous TVA experience. The maintenance cost was calculated as a percentage of the total direct investment based on previous TVA experience.

The capital charges portion of the indirect costs consists of: (1) depreciation, interim replacements, and insurance, which are based on the total depreciable investment; and (2) the average costs of capital and taxes, which are based on the total capital investment. The overhead charges consists of: (1) plant overhead, which is based on a percentage of the cost of operating labor and supervision, maintenance, and analyses; and (2) administrative overheads, which are based on a percentage of operating labor and supervision. The methods by which these capital charges are determined were described previously in the Economic Premises section.

The annual revenue requirements are the sum of the direct and the indirect costs. Equivalent unit revenue requirements in mills/kWh can then be obtained by dividing by the on-stream time of 7000 hr/yr and the plant net MW rating.

RESULTS

Both the annual revenue requirements and the equivalent unit revenue requirements were calculated for each of the seven NO_x FGT alternatives based on the design and economic premises of this study. A detailed breakdown of these revenue requirements for each process is provided on the following pages.

Asahi SO_x-NO_x Process

The total annual revenue requirement for the Asahi process is approximately $39.8M as shown in Table 46. This corresponds to a unit revenue requirement of 12.63 mills/kWh. Direct costs make up $22.0M or slightly more than 55% of the total. Indirect costs, primarily capital charges, account for the remaining 45.0%. The various capital charges of $15.1M represent 37.8% of the total annual revenue requirements. Although limestone is the largest in quantity, the major annual raw material costs are for sodium carbonate and EDTA, $3.55M and $1.66M respectively. Major conversion costs include fuel oil for cracking and flue gas reheat at $4.73M, electricity (primarily for the flue gas blowers) at $4.3M, and plant maintenance at $4.0M.

TABLE 46. ASAHI SO_x-NO_x PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
onsite solids disposal, 90% NO_x removal, 99% SO_2 removal)

	Annual quantity	Unit cost, $	Total annual cost, $	% of annual revenue requirement
Direct Costs				
Raw materials				
Limestone	113,000 tons	7.40/ton	836,200	2.10
$FeSO_4 \cdot 7H_2O$	840 tons	86.50/ton	72,700	0.18
$Na_4 \cdot EDTA$	2,240 tons	743.00/ton	1,664,300	4.18
Na_2CO_3	34,500 tons	103.00/ton	3,553,500	8.92
K_2SO_4	1,100 tons	144.00/ton	158,400	0.40
H_2SO_4	2,200 tons	54.00/ton	118,800	0.30
Total raw materials cost			6,403,900	16.08
Conversion costs				
Operating labor and supervision				
Plant	46,428 man-hr	12.50/man-hr	580,400	1.46
Solids disposal equipment	29,120 man-hr	17.00/man-hr	495,000	1.24
Utilities				
Fuel oil	10,500,000 gal	0.45/gal	4,725,000	11.86
Steam	663,000 MBtu	2.00/MBtu	1,326,000	3.33
Process water	590,700 kgal	0.12/kgal	70,900	0.18
Electricity	147,000,000 kWh	0.029/kWh	4,263,000	10.70
Landfill operation				
Trucks (fuel and maintenance)	198,200 tons	0.06/ton	11,900	0.03
Earthmoving equipment (fuel and maintenance)	198,200 tons	0.16/ton	31,700	0.08
Maintenance				
Labor and material			3,976,800	9.99
Analyses	8,760 man-hr	17.00/man-hr	148,900	0.37
Total conversion costs			15,629,600	39.24
Total direct costs			22,033,500	55.32
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			6,039,100	15.16
Average cost of capital and taxes at 8.6% of total capital investment			9,022,200	22.66
Overheads				
Plant, 50% of conversion costs less utilities			2,622,400	6.59
Administrative, 10% of operating labor			107,500	0.27
Total indirect costs			17,791,200	44.68
Total annual revenue requirements			39,824,700	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input
Equivalent unit revenue requirements	12.63	26.55	1.26

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a. Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Investment and revenue requirements for disposal of fly ash excluded.
Total direct investment, $56,811,000; total depreciable investment, $100,652,000; and total capital investment, $104,909,000.

IHI SO_x-NO_x Process

The annual revenue requirement for the IHI process is slightly less than $58.6M as shown in Table 47. This corresponds to a unit revenue requirement of about 19.82 mills/kWh. Annual direct costs are $25.5M, about 43.5% of the total annual revenue requirements. Indirect costs, primarily capital charges, account for the remaining 56.5%.

In contrast to the other wet, simultaneous SO_x-NO_x processes, the raw material costs for the IHI process are $1.8M, a relatively insignificant 3.1% of the annual revenue requirements. Although large quantities of some raw materials are used, they are relatively inexpensive materials such as limestone. The major direct costs are conversion costs, primarily electricity although maintenance and steam consumption also add significant amounts. Most of this electrical charge of $13.1M is consumed in generation of ozone. The major annual costs associated with the IHI process are the two capital charges (depreciation, etc., and average cost of capital) which together account for more than 50% of the total annual revenue requirement.

Moretana Calcium SO_x-NO_x Process

The total annual revenue requirement for the Moretana Calcium process is about $38.1M as shown in Table 48. This corresponds to an average unit revenue requirement of 11.73 mills/kWh. Annual direct costs for raw materials, labor, utilities, and maintenance are $23.6M or 62.1% of the total annual revenue requirements. Indirect costs, primarily for capital charges, account for the remaining 37.9%.

Although the annual direct cost is similar to that of the other wet, SO_x-NO_x processes, the raw material costs of $12.9M is significantly higher than the other processes evaluated. Limestone, the primary raw material in terms of quantity, is relatively cheap, but the Moretana Calcium process also consumes several expensive raw materials. The annual cost of hydrogen chloride (HCl) (used to generate ClO_2) and several proprietary additives contribute $4.3M and $4.1M, respectively, to the annual revenue requirements. The conversion costs, which are lower for the Moretana Calcium process than for the other wet processes, are about $10.7M, including electrical and maintenance costs of $5.5M and $3.0M respectively. The capital charges of nearly $12.5M represent 32.9% of the total annual revenue requirements.

Hitachi Zosen NO_x-Only Process

The total annual revenue requirement for the Hitachi Zosen process is approximately $12.2M as shown in Table 49. This is equivalent to a unit revenue requirement of about 3.61 mills/kWh. The direct costs, primarily raw material costs, constitute about $8.8M annually, or 71.8% of the total annual revenue requirements. Indirect costs, composed of capital charges and overheads, form 30% of the annual revenue requirement or about $3.7M. A credit for spent catalyst disposal of $0.26M, or 2.1% of the annual revenue requirement, is applied.

TABLE 47. IHI SO_x-NO_x PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
onsite solids disposal, 90% NO_x removal, 90% SO_2 removal)

	Annual quantity	Unit cost, $	Total annual cost, $	% of annual revenue requirement
Direct Costs				
Raw materials				
Limestone	114,400 tons	7.00/ton	800,800	1.37
H_2SO_4	770 tons	54.00/ton	41,600	0.07
Powdered limestone	11,800 tons	32.00/ton	377,600	0.65
NaCl	1,160 tons	25.80/ton	29,900	0.05
CuCl	77 tons	2,105.00/ton	162,100	0.28
NaOH	2,200 tons	187.00/ton	411,400	0.70
Total raw materials cost			1,823,400	3.12
Conversion costs				
Operating labor and supervision				
Plant	37,668 man-hr	12.50/man-hr	470,900	0.80
Solids disposal equipment	32,000 man-hr	17.00/man-hr	544,000	0.93
Utilities				
Fuel oil	1,102,000 gal	0.45/gal	495,900	0.85
Steam	1,463,000 MBtu	2.00/MBtu	2,926,000	5.00
Process water	853,700 kgal	0.10/kgal	85,400	0.15
Electricity	451,500,000 kWh	0.029/kWh	13,093,500	22.35
Landfill operation				
Trucks (fuel and maintenance)	235,800 tons	0.06/ton	14,100	0.02
Earthmoving equipment (fuel and maintenance)	235,800 tons	0.16/ton	37,700	0.06
Maintenance				
Labor and material			5,889,000	10.05
Analyses	6,570 man-hr	17.00/man-hr	111,700	0.19
Total conversion costs			23,668,200	40.40
Total direct costs			25,491,600	43.52
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			11,940,700	20.39
Average cost of capital and taxes at 8.6% of total capital investment			17,511,500	29.89
Overheads				
Plant, 50% of conversion costs less utilities			3,533,700	6.03
Administrative, 10% of operating labor			101,500	0.17
Total indirect costs			33,087,400	56.48
Total annual revenue requirements			58,579,000	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input
Equivalent unit revenue requirements	19.82	39.05	1.86

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[a] Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Investment and revenue requirements for disposal of fly ash excluded.
Total direct investment, $117,370,000; total depreciable investment, $199,011,000; and total capital investment, $203,622,000.

TABLE 48. MORETANA CALCIUM PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
onsite solids disposal, 90% NO_x removal, 95% SO_2 removal)

	Annual quantity	Unit cost, $	Total annual cost, $	% of annual revenue requirement
Direct Costs				
Raw materials				
Limestone	172,000 tons	7.00/ton	1,204,000	3.16
Slaked lime	14,000 tons	60.00/ton	840,000	2.21
CuCl	56 tons	2,105.00/ton	117,900	0.31
HCl	52,900 tons	81.00/ton	4,284,900	11.26
$(NH_4)_2SO_4$	28,250 tons	82.00/ton	2,316,500	6.09
Proprietary additives	2,380 tons		4,116,200	10.81
Total raw materials cost			12,879,500	33.84
Conversion costs				
Operating labor and supervision	46,428 man-hr	12.50/man-hr	580,400	1.52
Utilities				
Steam	736,100 MBtu	2.00/MBtu	1,472,200	3.87
Process water	365,300 kgal	0.12/kgal	43,800	0.11
Electricity	189,280,000 kWh	0.029/kWh	5,489,100	14.42
Maintenance				
Labor and material			3,005,900	7.90
Analyses	8,760 man-hr	17.00/man-hr	148,900	0.39
Total conversion costs			10,740,300	28.21
Total direct costs			23,619,800	62.05
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			4,954,500	13.02
Average cost of capital and taxes at 8.6% of total capital investment			7,565,200	19.87
Overheads				
Plant, 50% of conversion costs less utilities			1,867,600	4.91
Administrative, 10% of operating labor			58,000	0.15
Total indirect costs			14,445,300	37.95
Total annual revenue requirements			38,065,100	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input
Equivalent unit revenue requirements	11.73	25.38	1.21

a. Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Investment and revenue requirements for disposal of fly ash excluded.
Total direct investment, $46,691,000; total depreciable investment, $82,575,000; and total capital investment, $87,967,000.

TABLE 49. HITACHI ZOSEN PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal; 90% NO_x removal)

	Annual quantity	Unit cost, $	Total annual cost, $	% of annual revenue requirement
Direct Costs				
Raw materials				
NH_3	6,395 short tons	150.00/short ton	959,300	7.85
Catalyst			6,880,000	56.34
Total raw materials costs			7,839,300	64.19
Conversion costs				
Operating labor and supervision	8,760 man-hr	12.50/man-hr	109,500	0.90
Utilities				
Steam	8,096 MBtu	2.00/MBtu	16,200	0.13
Electricity	15,796,000 kWh	0.029/kWh	458,100	3.75
Heat credit	100,310 MBtu	2.00/MBtu	(200,600)	(1.64)
Maintenance				
Labor and material			497,900	4.07
Analyses	2,920 man-hr	17.00/man-hr	49,600	0.41
Total conversion costs			930,700	7.62
Total direct costs			8,770,000	71.81
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			1,358,800	11.13
Average cost of capital and taxes at 8.6% of total capital investment			2,004,400	16.41
Overheads				
Plant, 50% of conversion costs less utilities			328,500	2.69
Administrative, 10% of operating labor			11,000	0.09
Total indiiect costs			3,702,700	30.32
Spent cataylst disposal			(260,000)	(2.13)
Total average annual revenue requirements			12,212,700	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input
Equivalent unit revenue requirements for NO_x removal	3.61	8.14	0.39

a. Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Investment and revenue requirement for disposal of fly ash excluded.
Total direct investment, $12,447,000; total depreciable investment, $22,647,000; and total capital investment, $23,307,000.

As in all the dry, NO_x-only processes, NH_3 is the raw material used in the largest quantity. However, the major raw material cost is for catalyst, i.e., about $6.9M annually. The catalyst cost represents approximately 56.3% of the annual revenue requirements and is the largest single item in the annual revenue requirements. The major conversion costs are maintenance at almost $0.50M annually and electricity (primarily for the flue gas blowers) at $0.46M. A heat credit valued at $0.2M is applied based on the heat of reaction from the reduction of NO_x with NH_3 and the heat released from the combination of excess NH_3 with oxygen, less the preheat required for the NH_3 and air from the NH_3 injection system. The net heat released was considered to be recovered at the air heater.

The capital charges account for about 27.5% of the total annual revenue requirements, i.e., about $3.4M. Overheads make up 2.8% or $0.3M of the annual revenue requirements. The $0.26M credit for spent catalyst disposal is based on the scrap metal value of the catalyst base support.

Kurabo Knorca NO_x-Only Process

The total annual revenue requirement for the Kurabo Knorca process is approximately $9.3M as shown in Table 50. This is equivalent to a unit revenue requirement of 2.77 mills/kWh. The direct costs, composed of raw material and conversion costs, constitute about $6.0M annually or 63.9% of the total annual revenue requirements. Indirect costs, composed of capital charges and overheads, are 36.0% of the annual revenue requirements or about $3.4M. The spent catalyst disposal cost of $6k is less than 0.1% of the annual revenue requirement.

NH_3 is the raw material used in the largest quantity. However, the major raw material cost is for catalyst, i.e., about $4.3M annually. The catalyst cost represents approximately 46.2% of the annual revenue requirements. The major conversion costs are plant maintenance at $0.4M annually and electricity (primarily for flue gas blowers) at about $0.5M. A heat credit valued at $0.2M is applied based on the heat of reaction from the reduction of NO_x with NH_3 plus the heat released from the combination of excess NH_3 with oxygen, less the preheat required for the NH_3 and air from the NH_3 injection system. The net heat released was considered to be recovered in the air heater.

The capital charges are a major annual expense, accounting for $3.1M, or about 32.7% of the total annual revenue requirements. Overheads make up $0.3M or 3.3% of the annual revenue requirements. The spent catalyst disposal cost, based on disposal cost for TiO_2, the catalyst carrier, is estimated at $6k.

UOP SFGT-N, NO_x-Only Process

The total annual revenue requirement for the UOP SFGT-N process is approximately $7.2M as shown in Table 51. This is equivalent to a unit revenue requirement of 2.13 mills/kWh. The direct costs, comprising raw material and conversion costs, are about $4.2M, 58.8% of the total annual revenue requirements. Indirect costs, comprising capital charges and

TABLE 50. KURABO KNORCA PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
90% NO_x removal)

	Annual quantity	Unit cost, $	Total annual cost, $	% of annual revenue requirement
Direct Costs				
Raw materials				
NH_3	5,365 short tons	150.00/short ton	804,800	8.62
Catalyst	407 short tons	10,605.00/short ton	4,316,000	46.20
Total raw materials costs			5,120,800	54.82
Conversion costs				
Operating labor and supervision	8,760 man-hr	12.50/man-hr	109,500	1.17
Utilities				
Steam	5,978 MBtu	2.00/MBtu	12,000	0.13
Electricity	15,822,000 kWh	0.029/kWh	458,800	4.91
Heat credit	108,400 MBtu	2.00/MBtu	(216,800)	(2.32)
Maintenance				
Labor and material			436,000	4.67
Analyses	2,920 man-hr	17.00/man-hr	49,600	0.53
Total conversion costs			849,100	9.09
Total direct costs			5,969,900	63.91
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			1,234,400	13.21
Average cost of capital and taxes at 8.6% of total capital investment			1,822,300	19.51
Overheads				
Plant, 50% of conversion costs less utilities			297,600	3.19
Administrative, 10% of operating labor			11,000	0.12
Total indirect costs			3,365,300	36.03
Spent catalyst disposal			6,000	0.06
Total annual revenue requirements			9,341,200	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input
Equivalent unit revenue requirements for NO_x removal	2.77	6.23	0.30

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[a] Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Investment and revenue requirement for disposal of fly ash excluded.
Total direct investment, $10,899,000; total depreciable investment, $20,574,000; and total capital investment, $21,190,000.

TABLE 51. UOP SFGT-N, NO_x-ONLY PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
90% NO_x removal)

	Annual quantity	Unit cost, $	Total annual cost, $	% of annual revenue requirement
Direct Costs				
Raw materials				
NH_3	6,395 short tons	150.00/short ton	959,300	13.35
Catalyst			2,449,000	34.07
Total raw materials costs			3,408,300	47.42
Conversion costs				
Operating labor and supervision	8,760 man-hr	12.50/man-hr	109,500	1.52
Utilities				
Steam	7,140 MBtu	2.00/MBtu	14,300	0.20
Electricity	15,796,000 kWh	0.029/kWh	458,100	6.38
Heat credit	103,110 MBtu	2.00/MBtu	(206,200)	(2.87)
Maintenance				
Labor and material			393,800	5.48
Analyses	2,920 man-hr	17.00/man-hr	49,600	0.69
Total conversion costs			819,100	11.40
Total direct costs			4,227,400	58.82
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			1,085,100	15.10
Average cost of capital and taxes at 8.6% of total capital investment			1,584,400	22.05
Overheads				
Plant, 50% of conversion costs less utilities			276,500	3.85
Administrative, 10% of operating labor			11,000	0.15
Total indirect costs			2,957,000	41.15
Spent catalyst disposal			2,200	0.03
Total annual revenue requirements			7,186,600	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input
Equivalent unit revenue requirements for NO_x removal	2.13	4.79	0.23

a. Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Investment and revenue requirement for disposal of fly ash excluded.
Total direct investment, $9,844,000; total depreciable investment, $18,085,000; and total capital investment, $18,423,000.

overheads, are 41.2% of the annual revenue requirement or about \$3.0M. The spent catalyst disposal cost is \$2k, 0.03% of the annual revenue requirement.

NH_3 is the raw material used in the largest quantity. The cost of NH_3 is about 13.3% of the total annual revenue requirement, i.e., almost \$0.96M annually. The catalyst cost of \$2.4M, the largest individual expense, represents approximately 34.1% of the annual revenue requirements. Major conversion costs are plant maintenance at \$0.4M and electricity (primarily for flue gas blowers) at \$0.5M. A heat credit valued at about \$0.2M was applied based on the heat of reaction from the reduction of NO_x with NH_3 plus the heat released from the combination of excess NH_3 with oxygen, less the preheat required for the NH_3 and air from the NH_3 injection system. The net heat released was considered to be recovered in the air heater.

The capital charges account for about 37.1% of the total annual revenue requirements, i.e., about \$2.7M. Overheads are 4.0% or \$0.29M of the annual revenue requirements. The spent catalyst disposal cost, based on disposal cost for alumina, the catalyst carrier, is estimated to be \$2k.

UOP SFGT-SN, SO_x-NO_x Process

The total annual revenue requirements for the UOP SFGT-SN process for simultaneous SO_x-NO_x removal are approximately \$22.5M as shown in Table 52. This is equivalent to a unit revenue requirement of 6.64 mills/kWh. The direct costs, comprising raw material and conversion costs are 67.0% of the total annual revenue requirements or about \$15M. Indirect costs comprising capital charges and overheads, are 49.3% of the total annual revenue requirements. The spent catalyst disposal cost is \$6k, only 0.03% of the annual revenue requirement. A credit for byproduct sulfuric acid sales revenue of \$3.66M or 16.3% of the total annual revenue requirements is applied.

Naphtha is the raw material used in the largest quantity and represents a raw material cost of \$4.2M. Catalyst cost is the major raw material cost of \$6.6M. Major conversion costs are: (1) electricity costs (primarily for flue gas blowers, sulfuric acid plant, and quench circulating pumps) of about \$2.0M; (2) plant maintenance cost of \$1.5M; and (3) naphtha for fuel costs of almost \$1.1M. A heat credit valued at \$2.2M is applied. The net heat credit was based on the following:

1. Credit for heat reductions (oxidation of Cu, sulfation of CuO, and reduction of $CuSO_4$ and CuO).
2. Credit for heat of reactions for NO_x reduction with NH_3 and combination of excess NH_3 with oxygen.
3. Credit for combustion of purged gas.
4. Credit for vented gas combustion.
5. Credit for dewpoint suppression as a result of SO_3 removal.
6. Debit for NH_3, air, and steam superheat.

TABLE 52. UOP SFGT-SN, SO_x-NO_x PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
90% NO_x removal, 90% SO_2 removal)

	Annual quantity	Unit cost, $	Total annual cost, $	% of annual revenue requirement
Direct Costs				
Raw materials				
NH_3	6,395 short tons	150.00/short ton	959,300	4.27
Naphtha	8,400,000 gal	0.50/gal	4,200,000	18.69
Catalyst			6,648,000	29.59
Steam-naphtha reformer catalyst			100,000	0.45
Total raw materials cost			11,907,300	53.00
Conversion costs				
Operating labor and supervision	29,200 man-hr	12.50/man-hr	365,000	1.62
Utilities				
Naphtha	2,100,000 gal	0.50/gal	1,050,000	4.67
Steam	112,308 MBtu	2.00/MBtu	224,600	1.00
Process water	2,134,000 kgal	0.06/kgal	128,000	0.57
Electricity	70,550,000 kWh	0.029/kWh	2,046,000	9.11
Heat credit	1,121,680 MBtu	-2.00/MBtu	(2,243,400)	(9.99)
Maintenance				
Labor and material			1,492,900	6.65
Analyses	4,380 man-hr	17.00/man-hr	74,500	0.33
Total conversion costs			3,137,600	13.96
Total direct costs			15,044,900	66.96
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			3,931,300	17.50
Average cost of capital and taxes at 8.6% of total capital investment			5,775,100	25.71
Overheads				
Plant, 50% of conversion costs less utilities			966,200	4.30
Administrative, 10% of operating labor			36,500	0.16
Marketing, 10% of sales revenue			366,000	1.63
Total indirect costs			11,075,100	49.30
Spent catalyst disposal			6,000	0.03
Gross average annual revenue requirements			26,126,000	116.29
Byproduct Sales Revenue				
H_2SO_4	122,000 short tons	-30.00/short ton	(3,660,000)	(16.29)
Total byproduct sales revenue			(3,660,000)	
Total annual revenue requirements			22,466,000	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input
Equivalent unit revenue requirements for SO_2 and NO_x removal	6.64	14.98	0.71

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a. Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Investment and revenue requirement for disposal of fly ash excluded.
Total direct investment, $37,322,000; total depreciable investment, $65,521,000; and total capital investment, $67,153,000.

The capital charges are the major expense and account for 43.2%, i.e., about \$9.7M, of the total annual revenue requirement. Overheads are 6.1% or about \$1.4M, of the total annual revenue requirements. The spent catalyst disposal cost, based on disposal cost for alumina, the catalyst carrier, is estimated to be \$6k.

The \$3.66M credit for sulfuric acid sales revenue is based on the annual sale of about 122,000 short tons of 100% H_2SO_4.

Limestone FGD

The annual revenue requirement for the limestone FGD process with 90% SO_2 removal, which is used in conjunction with each of the dry NO_x-only FGT processes for overall system comparability with SO_x-NO_x processes, was estimated to be \$14,651,300 (17).

ESP

The annual revenue requirements of the various ESP's were estimated by the same procedures as given for the NO_x FGT processes and are summarized below.

1. The annual revenue requirement for four 99.5% efficient cold ESP's, to be used with the Hitachi Zosen and UOP SFGT-N processes, was estimated to be \$2,220,000. The electricity consumption was based on 2 watts/ft^2 of collecting area.

2. The annual revenue requirement for four 99.5% efficient cold ESP's, to be used with the UOP SFGT-SN process, was estimated to be \$2,983,500. The electricity consumption was based on 2 watts/ft^2 of collecting area.

3. The annual revenue requirement for four 98% efficient hot ESP's, to be used with the Kurabo Knorca process, was estimated to be \$2,687,300. The electricity usage was based on 4 watts/ft^2 of collecting area.

4. The annual revenue requirement for one 94% efficient cold ESP, to be used with the Moretana Calcium process, was estimated to be \$1,497,200. The electricity usage was based on 2 watts/ft^2 of collecting area.

Overall Comparison

The estimated total capital investment and annual revenue requirements for each of the NO_x FGT alternatives based on the previously described design and economic premises are shown in Tables 53 and 54 respectively. The total capital investments range from \$79.6M (\$165/kW) for the UOP SFGT-N system to \$203.6M (\$482/kW) for the IHI SO_x-NO_x system.

The three dry NO_x-only processes, UOP, Kurabo, and Hitachi Zosen, have total capital investments of $18.4M, $21.2M, and $23.3M respectively. Including a total capital investment of $50.4M for the limestone FGD system, and either $10.8M or $12.1M for an ESP, depending upon the NO_x process, the total capital investment for comparison of SO_x-NO_x-PM systems employing dry NO_x-only processes combined with limestone FGD and ESP is $79.6M, $83.7M, and $84.5M for the UOP, Kurabo, and Hitachi Zosen systems respectively.

The UOP SFGT-SN process, the only dry SO_x-NO_x process, has a total capital investment cost of about $67.2M. With the addition of $14.6M for PM removal with ESP, the total capital investment for the SO_x-NO_x-PM system is $81.8M (i.e., $169/kW). This investment is 2.4% greater than for the UOP SFGT-N system (least costly SCR-FGD system), but 3.5% less than for the Hitachi Zosen system (most costly SCR-FGD system).

TABLE 53. TOTAL CAPITAL INVESTMENT

FOR THE ALTERNATIVE SO_x-NO_x-PM SYSTEMS

	Total capital investment, M$[a]				$/kW	$/aft^3/min
Process	FGT	FGD	ESP	Total	Total	Total
UOP SFGT-N	18.4	50.4	10.8	79.6	165	37.1
UOP SFGT-SN[b]	67.2	-	14.6	81.8	169	38.1
Kurabo Knorca	21.2	50.4	12.1	83.7	174	39.0
Hitachi Zosen	23.3	50.4	10.8	84.5	175	39.4
Moretana Calcium	88.0	-	7.2	95.2	205	44.4
Asahi	104.9	-	-	104.9	233	48.9
IHI	203.6	-	-	203.6	482	94.9

a. Basis
Midwest plant location represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979.
Stack gas reheat to 175°F by indirect steam reheat.
Minimum in-process storage; only pumps are spared.
Disposal pond located 1 mile from power plant.
Investment requirements for fly ash disposal excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

b. Based on hydrogen production from naphtha and H_2SO_4 byproduct from SO_2.

The total capital investments of the wet SO_x-NO_x processes are substantially higher and more divergent than those of the dry SCR-FGD systems. With an ESP cost of $7.2M, the Moretana Calcium system at $95.2M ($205/kW) is approximately 24% higher than the least capital-intensive dry, SCR-FGD system (UOP SFGT-N) while the Asahi system at $104.9M ($233/kW) is an additional 14% higher than the Moretana Calcium system. (It should be noted that to obtain 90% NO_x reduction, Moretana Calcium and Asahi processes achieve 95% and 99% SO_2 removal which is greater than the 90% SO_2 removal of the other alternative FGT processes.) The IHI system has a total capital investment of \$203.6M (\$482/kW), approximately double that of the Asahi system and almost 3 times that of the least capital-intensive dry systems.

The total annual revenue requirements for SO_x-NO_x-PM control ranged from 7.13 mills/kWh for the cheapest dry process (UOP SFGT-N) to more than 19.8 mills/kWh for the IHI process. When combined with total annual revenue requirements for the limestone FGD system and an ESP, the revenue requirements for the three SCR-FGD systems ranged from 7.13 mills/kWh for the UOP SFGT-N system to 8.60 mills/kWh for the Hitachi Zosen system. The revenue requirement for the UOP SFGT-SN process, the only dry SO_x-NO_x system, was 6.6 mills/kWh (including a byproduct sales credit). With the addition of the ESP, the revenue requirement for this SO_x-NO_x-PM system becomes 7.53 mills/kWh. This revenue requirement is only 6% higher than the least expensive SCR-FGD system. Without the credit for sale of byproduct H_2SO_4, the revenue requirement for the UOP SFGT-SN process plus ESP becomes 8.60 mills/kWh. This is 21% higher than the least costly SCR-FGD system.

TABLE 54. TOTAL ANNUAL REVENUE REQUIREMENTS

FOR THE ALTERNATIVE SO_x-NO_x-PM SYSTEMS

	Annual revenue requirements, M\$[a]				Equivalent revenue requirements, mills/kWh
Process	FGT	FGD	ESP	Total	Total
UOP SFGT-N	7.2	14.7	2.2	24.1	7.13
UOP SFGT-SN[b]	22.5	-	3.0	25.5	7.53
Kurabo Knorca	9.3	14.7	2.7	26.7	7.91
Hitachi Zosen	12.2	14.7	2.2	29.1	8.60
Moretana Calcium	38.1	-	1.5	39.6	12.20
Asahi	39.8	-	-	39.8	12.63
IHI	58.6	-	-	58.6	19.82

a. Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Investment and revenue requirement for disposal of fly ash excluded.

b. Based on hydrogen production from naphtha and H_2SO_4 byproduct from SO_2

The wet NO_x FGT systems are significantly more expensive to operate than the dry systems. The Asahi systems and Moretana Calcium systems at 12.6 mills/kWh and 12.2 mills/kWh, respectively, are the cheapest of the wet processes but are still nearly twice as expensive to operate as the less expensive dry NO_x-only - limestone FGD and simultaneous SO_x-NO_x systems. In addition, the IHI system at 19.8 mills/kWh is significantly higher. Thus it would appear from the total annual revenue requirements that the wet SO_x-NO_x processes cannot economically compete with the dry SCR-FGD or SO_x-NO_x processes under the premises used for this study. Under different premises, however, it may be possible that these wet SO_x-NO_x processes might be feasible, e.g., with retrofit if auxiliary heat were required with the dry FGT alternatives, or if various byproduct credit situations were considered.

Based on an accuracy range of -20% to +40%, the resulting range of estimated capital investment and annual revenue requirement for the SO_x-NO_x-PM control systems is given in Table 55.

Energy Requirements for the Alternative NO_x FGT Systems

Because of increasing concern about the energy requirements of various processes, evaluation of energy costs is important. A comparison of the energy requirements for the various alternative NO_x FGT systems is shown in Table 56. Fuel, steam, and heat credits are listed in equivalent MBtu/hr. The electrical consumption is shown as equivalent MBtu/hr. The total equivalent energy consumption is listed as a percentage of the boiler capacity. The energy consumption for the limestone FGD system and the various ESP's has been included.

The dry systems have the lowest energy requirement; energy equivalent to 4% of the boiler capacity is consumed. The UOP SFGT-SN process requires over 3% of the boiler capacity. The actual equivalent energy consumption, however, has been reduced by about 160 MBtu/hr through the application of the various heat credits claimed by UOP. Where applicable, similar heat credits have been given to the other NO_x FGT processes. Combined with the energy for an ESP, the energy for the UOP SFGT-SN system totals 3.6% of the boiler capacity.

The dry NO_x-only processes have low energy requirements, much less than 1% of boiler capacity in all three cases. Combined with the energy costs for the limestone FGD system and the ESP, energy costs for the SCR-FGD systems are still relatively low, ranging from 3.7 to 3.9% of boiler capacity.

The energy requirements for the wet systems range from 365 MBtu/hr (8.1% of the boiler capacity) for the Moretana Calcium system, with an ESP, to 836 MBtu/hr for the IHI system (18.6% of the boiler capacity). Thus the wet simultaneous systems are 2 to nearly 5 times as energy intensive as the dry NO_x FGT alternatives. In addition, consuming more than 10% of the new boiler capacity would severely affect new coal-fired boiler economics.

TABLE 55. ACCURACY RANGE OF THE ESTIMATED CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENT FOR THE SO_x-NO_x-PM SYSTEMS

(Based on a -20% to +40% range)

Process	Capital investment, $/kW			Unit revenue requirements, mills/kWh		
	Low	Base	High	Low	Base	High
UOP SFGT-N	132	165	231	5.70	7.13	9.98
UOP SFGT-SN[a]	135	169	237	6.02	7.53	10.54
Kurabo Knorca	139	174	244	6.33	7.91	11.07
Hitachi Zosen	140	175	245	6.88	8.60	12.04
Moretana Calcium	164	205	287	9.76	12.20	17.08
Asahi	186	233	326	10.10	12.63	17.68
IHI	386	482	675	15.86	19.82	27.75

a. Based on hydrogen production from naphtha and H_2SO_4 byproduct from SO_2.

TABLE 56. COMPARISON OF ENERGY REQUIREMENTS FOR THE VARIOUS NO_x FGT PROCESSES[a]

Process	Fuel, MBtu/hr	Steam, MBtu/hr	Electricity, MBtu/hr	Heat credit, MBtu/hr	Total equivalent energy consumption,[b] % of boiler capacity
Limestone FGD	-	78	74	-	3
Cold ESP (99.5% efficient)	-	-	8	-	0.2
Cold ESP (99.5% efficient - UOP SFGT-SN)	-	-	13	-	0.3
Cold ESP (94% efficient)	-	-	4	-	0.1
Hot ESP (98% efficient)	-	-	19	-	0.4
Asahi	216	105	189	-	11
IHI	23	232	581	-	19
Moretana Calcium	-	117	244	-	8
Moretana Calcium and ESP (94%)	-	117	248	-	8
Hitachi Zosen	-	1	20	14	0.2
Hitachi Zosen and FGD and ESP (99.5%)	-	79	102	14	4
Kurabo Knorca	-	1	20	16	0.1
Kurabo Knorca and FGD and ESP (98%)	-	79	113	16	4
UOP SFGT-N	-	1	20	15	0.2
UOP SFGT-N and FGD and ESP (99.5%)	-	79	102	15	4
UOP SFGT-SN	200	18	91	160	3
UOP SFGT-SN and ESP (99.5%)	200	18	104	160	4

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a. Does not include energy requirement represented by raw materials.

b. Based on a 500-MW boiler, a gross heat rate of 9,000 Btu/kWh for generation of electricity, and a boiler efficiency of 90% for generation of steam.

Wet Process Gas-Phase Oxidation Versus Absorption of NO_x

The IHI and Moretana SO_x-NO_x processes based on gas-phase oxidation were devised to avoid absorption of relatively insoluble NO and thus reduce the cost of the SO_x-NO_x absorption section. However, it appears that the choice of oxidant dictates whether oxidation is more economical than absorption. The ozone generating system cost is 38.3% of the total capital investment for the IHI process. In fact, the direct capital investment for the ozone generating system is nearly as much as the total capital investment of entire SCR-FGD systems. The investment for the chlorine dioxide system in the Moretana Calcium process is about 19% of the capital investment, a lesser but still significant cost, reflecting the relatively large capital investment of the IHI process. However, the Asahi process, an absorption-reduction-type process, has a higher capital investment than the Moretana Calcium process primarily as a result of the capital costs for systems to absorb the NO_x and to regenerate the sodium-based scrubbing solution.

The annual revenue requirements for both the Asahi and the Moretana Calcium systems are 12.6 mills/kWh and 12.2 mills/kWh, respectively, while IHI at 19.8 mills/kWh is significantly more expensive. The capital charges are primarily responsible for the differences in the annual revenue requirements, $29.5M for IHI versus $12.5M for Moretana Calcium and $15.1M for Asahi. Although the direct costs of each of these processes are essentially the same at $22.0M to $25.5M for each, the annual raw material costs are $1.8M for IHI, $6.4M for Asahi, and $12.9M for the Moretana Calcium process. Annual conversion costs vary from $10.7M for the Moretana Calcium process to $23.7M for IHI.

The use of an iron-chelating compound to assist the absorption of NO rather than oxidizing NO to NO_2 reduces capital investment because a less expensive nitrogen byproduct section is required. The Asahi process which uses $Fe^{+2}\cdot$EDTA converts most of the absorbed NO_x to molecular nitrogen and only relatively small amounts of complex nitrogen-sulfur compounds are formed.

The use of a gas-phase oxidant on the other hand results in the conversion of the absorbed NO_x to lesser amounts of molecular nitrogen and larger amounts of either nitrate salts or a combination of nitrate salts and complex nitrogen-sulfur compounds, depending on the gas-phase oxidant used. The use of ClO_2 as the gas-phase oxidant in the Moretana Calcium process results in 50% of the absorbed NO_x forming nitrate salts and the remaining 50% being converted to molecular nitrogen. The IHI process using ozone results in only 10-15% of the absorbed NO_x being converted to molecular nitrogen with most of the remainder forming complex nitrogen-sulfur compounds.

The capital investment costs for the nitrogen treatment sections reflect this inverse relationship between investment and the amount of NO_x reduced to molecular nitrogen in each process. The total capital investment of the nitrogen treatment section of the Asahi process, which forms the higher ratio of molecular nitrogen, is the lowest at slightly more than $1.17M, while in the IHI process, which converts only 10-20% of the absorbed NO_x to molecular nitrogen, the total capital investment for the nitrogen treatment section is about $5.15M.

Thus it appears that wet processes based on oxidation of NO_x with ozone (IHI) are not economically competitive with the other wet FGT processes primarily because of the costs involved in ozone generation. The use of ClO_2 as the oxidant, as in the Moretana Calcium process, results in a more favorable economic comparison with the absorption-reduction Asahi process. The Asahi process, which is about 14% higher in total capital investment than the Moretana Calcium process and similar to it in revenue requirements, cannot be definitely ruled out on the basis of this preliminary study. Under other premises--a higher NO_x concentration in the flue gas, for example--the economic positions of these two processes could be reversed.

Wet SO_x-NO_x Versus Dry SO_x-NO_x

From a comparison of total capital investments, the dry SO_x-NO_x system (UOP SFGT-SN), at about $81.8M ($169/kW), is much less expensive than any of the wet SO_x-NO_x systems, which range from 21% to 185% higher. The IHI system, with 185% higher capital costs than the UOP system does not appear economically competitive. Because of the preliminary nature of these estimates, however, it is impossible to declare either the Asahi or the Moretana Calcium system as economically unfeasible compared with UOP solely on the basis of total capital investment.

A comparison of the annual revenue requirements for UOP and the wet systems, however, indicates that the wet SO_x-NO_x FGT systems are not economically competitive. Annual revenue requirements for the UOP system are 7.53 mills/kWh with byproduct credit and 8.60 mills/kWh without this credit whereas those for the Asahi, Moretana Calcium, and IHI systems, are about 12.63, 12.20, and 19.82 mills/kWh respectively. With about 62% to 163% higher annual revenue requirements (42% to 130% without byproduct credit for UOP system) and about 21% to 185% higher capital investment, it would appear that the more expensive wet SO_x-NO_x systems would be eliminated from further contention for most applications. However, it would be difficult to extrapolate the result of these base case estimates to all potential conditions.

Another factor to consider is the relatively early status of development of the Asahi process. With the maturation of the Asahi technology it is possible that the various costs could be reduced to make the process competitive. However, it should also be pointed out that in most instances as the technology has been further refined, the costs have tended to increase rather than decrease.

Wet SO_x-NO_x Versus SCR-FGD

The total capital investments for the various combination SCR-FGD systems, which ranged from $79.6M ($165/kW) to $84.5M ($175/kW), were less than the investment required for the various wet SO_x-NO_x systems. The wet systems ranged from 24% to 192% higher than the UOP SFGT-N FGD system, which had a total capital investment of $79.6M. Based on comparison of capital investments it appears that the Asahi and IHI systems are not economically competitive when compared with the SCR-FGD systems under the premises of this study.

A comparison of the annual revenue requirements reinforces this view. The annual revenue requirements for the three SCR-FGD systems range from 7.13 mills/kWh for UOP to 8.60 mills/kWh for Hitachi Zosen, while those for the wet SO_x-NO_x systems range from 12.20 to 19.82 mills/kWh, 71 to 178% higher than the UOP SFGT-N FGD system. Thus it would appear that for most applications, the wet SO_x-NO_x systems (Asahi, IHI, and Moretana Calcium processes) are not economically competitive with the SCR-FGD systems.

Dry SO_x-NO_x Versus SCR-FGD

The total capital investment of \$81.8M (\$169/kW) for the UOP SO_x-NO_x system is between the investment for SCR-FGD combination systems, which range from \$79.6M to \$84.5M in total capital investment. Any difference is well below the accuracy of these capital investment estimates so that there is no significant difference for these two types of FGT within the premises of this study.

The annual revenue requirements comparison has similar results. The annual revenue requirements for the three SCR-FGD systems range from 7.13 mills/kWh for UOP to 8.60 mills/kWh for the Hitachi Zosen system while the revenue requirements for the UOP SO_x-NO_x system are 7.53 mills/kWh with byproduct credit and 8.60 mills/kWh without the credit. Thus, with the byproduct credit the annual revenue requirement for the dry SO_x-NO_x removal system is 6% higher than the UOP SFGT-N system. The other SCR-FGD systems range from 5% (for Kurabo) to 14% (for Hitachi Zosen) higher than the UOP SFGT-SN system. Without the byproduct credit, the UOP SFGT-SN system is 21% and 9% higher than the UOP SFGT-N and Kurabo systems, respectively, and the same as the Hitachi Zosen system. However, these results may vary under different cases, for example, if the annual catalyst replacement on these dry processes (which was a large contributing factor) were significantly different. Also, for the cost estimate of the UOP SFGT-SN process, hydrogen production from naphtha and conversion of $SO2$ to H_2SO_4 byproduct are assumed. However, if the future supply of naphtha became uncertain, it may be necessary to produce hydrogen by an alternative method, such as, coal gasification. In addition, if there is no market for $H2SO_4$ byproduct, SO_2 processing alternatives may be used, for example, production of elemental sulfur or liquid SO_2. Inclusion of alternative processing schemes for the UOP SFGT-SN process may significantly alter these economic evaluations.

The sensitivity of the dry processes to the cost of NH_3 has been the subject of some concern. In this study the annual NH_3 cost was less than 10% of the annual revenue requirement for two of the three dry NO_x FGT processes (Hitachi Zosen and Kurabo Knorca) and the NH_3 cost accounted for 13% of the annual revenue requirement for the remaining dry process (UOP SFGT-N). Thus, under the premises used in this study, the average annual revenue requirement may be expected to increase about 10% when the NH_3 cost is doubled.

Moving-Bed Reactor Versus Parallel Flow, Fixed-Bed Reactor

The capital investment and annual revenue requirements for the Kurabo Knorca moving-bed reactor process are \$43.9/kW and 2.77 mills/kWh, respectively, for only the NO_x removal portion. The capital investment and annual revenue requirements for the parallel flow, fixed-bed reactor processes are \$38.1/kW and 2.13 mills/kWh for UOP SFGT-N, and \$48.2/kW and 3.61 mills/kWh for Hitachi Zosen. Thus, the costs for the moving-bed reactor fall between the range of costs for the parallel flow, fixed-bed processes. Therefore, no significant distinction can be made based on this level of study between the moving-bed and parallel flow, fixed-bed processes. Actually, the major economic differences among the NO_x-only FGT processes are the catalyst costs. The annual catalyst cost, based on a 1-year-guaranteed life and in 1980 dollars, ranged from about \$2.5M for UOP to almost \$6.9M for Hitachi Zosen.

CONCLUSIONS AND RECOMMENDATIONS

CONCLUSIONS

Several conclusions may be derived from the data produced in this study. The major conclusions are listed below. It should be remembered that these conclusions apply to the results from our base case and premises involving a new, 500-MW coal-fired unit. The results and conclusions may vary with different cases, e.g., smaller or larger plant, lower or higher sulfur content in coal, different NO_x and SO_x removal efficiencies, or various credits for byproduct sales. Also, these results are based on currently available information and may have to be revised as future testing better defines the FGT technology and its impact on design and operation of downstream equipment.

Capital Investment

1. A dry SCR-FGD system (UOP SFGT-N) has the lowest capital investment. The IHI SO_x-NO_x removal process has the highest investment.

2. The wet SO_x-NO_x processes based on ozone as a gas-phase oxidant are prohibitively expensive for coal-fired units where NO_x concentrations are relatively high (e.g., 600 ppm).

3. The wet gas-phase oxidation process using ClO_2 (Moretana Calcium process) requires less investment than the other wet FGT processes. The investments for the Asahi and IHI processes were 14 and 135% higher, respectively, than the Moretana Calcium process.

4. The dry SO_x-NO_x simultaneous removal system (UOP SFGT-SN) has a lower investment than the wet, simultaneous SO_x-NO_x control systems. The investment ranges from 21% higher for the Moretana Calcium system to 185% higher for the IHI system, than the dry UOP SFGT-SN removal system.

5. The dry NO_x-only removal processes in combination with a limestone FGD unit have lower investments than the wet, simultaneous SO_x-NO_x removal systems. In comparison with the least expensive dry NO_x-only process and limestone FGD unit, the investment ranges from 24% (for Moretana Calcium) to 192% (for IHI) higher for the wet SO_x-NO_x removal systems.

6. The dry, simultaneous SO_x-NO_x control process has an essentially equivalent investment as the combination of dry NO_x-only removal process and limestone FGD unit. The UOP SFGT-SN process has a 2.4% greater investment than the least expensive SCR-FGD system (UOP SFGT-N) and 3.5% lower investment than the most expensive SCR-FGD system (Hitachi Zosen).

7. At this preliminary level of study, there is no significant difference in the investment among the dry NO_x-only FGT processes. In combination with the limestone FGD unit, the most expensive NO_x-only control process (Hitachi Zosen) has just a 6.0% higher investment than the least expensive process (UOP SFGT-N).

Annual Revenue Requirement

1. The UOP SFGT-N process has the lowest annual revenue requirement. The IHI SO_x-NO_x removal process has the highest revenue requirement.

2. The revenue requirement is similar for two of the wet processes while one process has substantially larger revenue requirements. The revenue requirement is 3.5% and 62% higher for the Asahi and IHI processes, respectively, than the Moretana Calcium process.

3. The revenue requirement is significantly less for the dry SO_x-NO_x removal system than the wet SO_x-NO_x control systems. The revenue requirement ranges from 62% greater for the Moretana Calcium system to 163% greater for the IHI system than for the dry UOP SFGT-SN system (with byproduct credit). In comparison to the UOP SFGT-SN system without byproduct credit, the wet systems have 42% to 130% higher revenue requirements.

4. The SCR-FGD systems have significantly lower revenue requirements than the wet SO_x-NO_x control systems. The revenue requirement ranges from 71% to 178% higher for the Moretana Calcium and IHI systems, respectively, than for the SCR-FGD system with the lowest revenue requirement (i.e., UOP).

5. The revenue requirement for dry SO_x-NO_x removal is in the range of revenue requirements for the SCR-FGD systems. With the byproduct credit the UOP SFGT-SN removal system has a 6% higher revenue requirement than UOP SFGT-N process plus FGD unit. The other SCR-FGD systems range from 5% (for Kurabo) to 14% (for Hitachi Zosen) higher in revenue requirements than the UOP SFGT-SN system. Without the byproduct credit, the UOP SFGT-SN system has 21% and 9% higher revenue requirement than the UOP SFGT-N and Kurabo SCR-FGD systems, respectively, but has an equal revenue requirement to the Hitachi Zosen system.

6. The Moretana Calcium process has the greatest annual raw material cost, i.e., almost \$12.9M. The IHI process has the lowest annual raw material cost of \$1.8M.

7. The least energy-intensive system is the UOP simultaneous SO_x-NO_x control system with a net energy consumption of 162 MBtu/hr with the heat credit (322 MBtu/hr without the heat credit). The IHI system is the most energy-intensive using 835.4 MBtu/hr.

8. For the dry NO_x-only FGT processes, the revenue requirement would be expected to increase about 10% as the NH_3 cost is doubled.

RECOMMENDATIONS

Additional preliminary- and definitive-level economic assessments of NO_x control systems should be performed. The results and conclusions of this NO_x FGT study are based on a preliminary-level estimate for the base case of treating flue gas from a new, 500-MW coal-fired power plant. It is very possible that the results would be different with variations from the abovementioned base case and thus extrapolations of these results to other conditions could lead to erroneous conclusions. Also, current and future development testing will better define the FGT technology and its impacts on downstream equipment, such as the air heater, FGD system, ESP or baghouse, and flue gas fan. In addition, combustion modification for NO_x emission control, which has received the primary attention in the United States and is considered the least expensive NO_x reduction system, has not been considered in this study. Therefore, additional economic evaluations, including FGT and combustion modifications, should be conducted. The FGT processes and combustion modification could be evaluated both independently and in combination with each other. In addition, these control techniques should be compared under several case variations to determine the sensitivity of these technologies to various design and economic considerations. Suggested items which should be varied include treatment capacity, sulfur composition in coal, total NO_x removal efficiency, raw material stoichiometric ratios, and byproduct sales revenue.

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16. ABSTRACT

The report gives results of a preliminary economic evaluation of seven flue gas treatment (FGT) processes for the removal of NOx from power plant flue gas. The base case was a new, 500-MW power plant burning 3.5% sulfur coal and emitting 600 ppm NOx in the flue gas. Total capital investments and annual revenue requirements for three dry NOx-only removal processes were $38-48/kW and 2.1-3.6 mills/kWh, respectively. A dry process to simultaneously remove both SOx and NOx had a capital cost of $139/kW and an annual revenue requirement of 6.6 mills/kWh. Wet SOx/NOx systems, which also remove particulates, had capital costs and annual revenue requirements of $205-482/kW and 12.2-19.8 mills/kWh, respectively. To permit comparisons between the alternative NOx control systems, the costs of both a limestone slurry flue gas desulfurization unit and an ESP were added to the cost of the dry NOx-only processes, and the cost of an ESP was added to the cost of the dry SOx/NOx processes. Total capital investments for these combined systems based on dry NOx removal (both NOx-only and SOx/NOx removal) were about $165-175/kW. Unit revenue requirements for the combined systems were 7.1-8.6 mills/kWh for the dry processes. For the base case, wet FGT processes do not appear to be economically competitive with either the dry NOx-only or SOx/NOx systems.

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Economic Analysis	Desulfurization	Stationary Sources	05C	07A,07D
Flue Gases		Flue Gas Treatment	21B	
Nitrogen Oxides			07B	
Sulfur Oxides				
Coal			21D	

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